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DEVELOPMENT OF A NEW GRAVITY SEDIMENTATION PROCESS FOR DEWATERING FLUE GAS CLEANING WASTES

by

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16. ABSTRACT The report gives results of a project to develop and test a novel system for dewatering flue gas cleaning (FGC) wastes at the pilot plant level. In this new system, the clarification and thickening functions are conducted in separate, but interconnected, pieces of equipment. The new system consists of a lamella clarifier and a conventional thickener that is smaller in diameter, but deeper (than the thickener/clarifier typically used to dewater FGC wastes), connected by a recycle stream between the two units to obtain a high degree of flexibility and control of operating conditions. Preliminary economic evaluation of this system indicates potential savings of 10 percent of the total capital costs and 6 percent of annual operating costs for the FGC waste management/disposal system. In pilot testing of this system, a completely new concept in thickener operation, known as the 'bang-bang' operation, evolved, in which the thickener underflow rate is set as low as possible without plugging the underflow lines. Periodically, the underflow rate is increased briefly to remove additional solids from the system at the concentration established by the (previously set) low underflow rate. This mode of operation appears to make it possible to maintain the maximum solids concentration in the underflow.

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Sedimentation		Stationary Sources	14G	13I
Dewatering			13H, 07A	
Flue Gases			21B	
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ABSTRACT

This final report presents the test results, the derived semi-empirical design routines, the process synthesis and developments, and the follow-up technology transfer efforts from May 1, 1976, through February, 1979, of the EPA sponsored research project "Development of a New Gravity Sedimentation Process for Dewatering Flue Gas Cleaning Wastes", EPA Grant No. R804531. In addition, the results from the shakedown and the final pilot plant testing of a novel dewatering system developed under this parent EPA contract are presented.

A completely new concept in thickener operation evolved from the pilot test results. This concept is known as 'bang-bang' operation. In this mode of operation, the thickener underflow rate is set as low as possible without plugging the underflow lines. Periodically, the underflow rate is stepped up for a short period in order to maintain an overall solids balance within the thickener. This mode of operation makes it possible to maintain a maximum underflow concentration. An additional feature of this mode of operation is that it is designed such that the functions of clarification and thickening can be effectively decoupled.

The pilot plant testing and follow-up analyses were planned and performed during the period following the completion of the original EPA grant to June 15, 1983 under sponsorship of a subsequent EPA grant (No. R806314).

A new, more efficient gravity sedimentation system for the dewatering of flue gas cleaning (FGC) wastes was conceived, designed, and tested on a laboratory pilot scale. The system consists of two units: a lamella clarifier and a smaller-diameter, deeper, conventional thickener. These units are connected in a unique way, so as to decouple the clarification and thickening functions of the system as much as possible. A recycle stream between the two units is used to obtain a greater degree of flexibility and control of operating conditions.

The objective of the parent EPA sponsored research project was to develop more economically feasible methods than those currently available for continuous treatment for FGC wastes. The objective was successfully accomplished in that a new gravity FGD dewatering system was developed having the potential of a maximum savings of ten percent of the total capital costs and six percent of operating costs for the entire FGC waste management/disposal system.

Under the original grant, equipment designs were made for a larger-scale pilot evaluation of the new dewatering system. The larger-scale pilot evaluation was carried out at TVA's pilot facility located at Paducah, Kentucky, under the subsequent EPA grant.

It was under the sponsorship of EPA Grant No. R806314 that the pilot plant evaluation was done. Support for the pilot plant equipment was provided by the Tennessee Valley Authority (TVA). The general concepts upon which the new dewatering system was based were shown to be generally valid. The system was proven to allow FGC wastes with high calcium sulfite content concentrated to a high solids content (50 wt.%) using reasonably sized equipment and no chemical additives.

It appears that nearly optimal use of the natural compressive forces of a settling sludge, like FGC waste, can be made when the new dewatering system is operated in the "bang-bang" mode.

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CONVERSION FACTORS

English/American Units	Metric Equivalent
Length:	
1 inch (in)	2.540 centimeters (cm)
1 foot (ft)	0.3048 meter (m)
Area:	
1 square ft (ft^2)	0.0929 square meter (m^2)
Volume:	
1 gallon	3.785 liters
Flow rate:	
1 gallon/minute (gpm)	3.785 liters/minute (l/min)
Pressure:	
1 pound per square inch (psi)	0.07031 kilogram per square centimeter (kg/cm^2)

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Section 1

SUMMARY

This is the final report on the EPA sponsored research projects "Dewatering Principles and Equipment Design Studies" and "Dewatering of Flue-Gas-Cleaning Wastes by Gravity Sedimentation-Pilot Evaluation". It covers the entire duration of the two projects --from May 1, 1976 through June 15, 1983. There were essentially four phases to the work. In Phase I, a preliminary evaluation was made to determine the direction of the program. To this end, limited sedimentation and filtration studies were made to disclose which offered the best potential for improving dewatering efficiencies. In Phase II, a detailed process synthesis and development effort was made. In this effort, fundamental concepts in gravity sedimentation and process dynamics were applied to synthesize a new system for dewatering flue gas desulfurization (FGD) wastes. In Phase III, a feasibility analysis was done to estimate the maximum savings possible with the use of the new dewatering system. Also, equipment designs, layouts, etc., for a larger-scale pilot evaluation of the new dewatering system were made. In Phase IV, a pilot-scale system was constructed, transported, and installed at TVA's pilot FGD facility located at the Shawnee Steam Plant near Paducah, Kentucky. Shakedown studies were done to reveal any possible need for equipment improvement. Equipment modifications were made in cooperation with TVA staff. Final pilot-plant tests were performed. The objectives of these tests were to confirm whether the clarification function of the unit could be decoupled from its thickening function, to

determine if the unit was capable of concentrating unoxidized FGD wastes to a high solids content, and to identify design limitations and make operational recommendations. The results of the pilot tests should promote the transfer of this new dewatering technology to industry, hopefully with its commercialization in the near future.

The work done under Phase I is discussed in Section 3. Settling data were collected showing that the FGD wastes tested exhibit a settling behavior similar to that of weakly flocculating materials. Assuming that FGD wastes do form weak flocs while settling, the flocs appear to have a low yield strength, and the gravitational force (the weight) exerted by overlying layers of solids in the settling medium compresses underlying layers. This means that the operating depth of the blanket of settling waste in the thickener is an important process parameter: the deeper the blanket of settling wastes at the bottom of the thickener, the higher the solids concentration in the underflow; or, the longer the residence time of solids in the compression zone of the thickener, the greater the dewatering. Batch and continuous settling data also showed that thickening rather than clarification was the controlling design factor: more settling area was required to thicken the FGD wastes than that needed for their clarification. These findings showed promise that the application of past studies in our laboratories could lead to a less expensive gravity sedimentation system for dewatering FGD wastes.

Limited filtration studies were done using a one square foot plate and frame filter press. The filtration data collected did not show much promise that our continued study of the filtration operation itself would yield much of an improvement over conventional filtration operations. Therefore, an alternate strategy was chosen: to improve the dewatering efficiency of the gravity sedimentation system, thereby increasing the solids concentration of the influent stream to the filters and lowering the filter dewatering load.

Phase II, the major thrust of the bench-scale part of this research project, is discussed in Sections 4-11. A new, more efficient gravity sedimentation system for the dewatering of FGD wastes was conceived, designed, and tested on a laboratory pilot scale. The system consists of two units: a laminar-type clarifier and a conventional thickener. These units are connected in a unique way, so as to decouple the clarification and the thickening functions of the system as much as possible. A recycle stream between the two units is used to obtain a greater degree of flexibility and control of operating conditions. The important operating parameters for the system were investigated by conducting a series of batch and continuous settling experiments. These parameters included: the location of the level of the solid/ liquid interface in the clarifier, the angle of inclination of the clarifier, and the flow rates of the feed and recycle streams.

The dewatering efficiency of the new system was evaluated on a laboratory pilot scale. The pilot thickener was six inches in diameter and about 15 feet in length. Tubular clarifiers were used. These were 1-3/4 inches in inside diameter, 6-1/2 feet in length, and inclined at an angle of 30°. The results of the pilot tests showed the new system to be effective in obtaining high underflow concentrations (30 to 60 percent) and a clear overflow at high throughput rates while requiring a smaller settling area compared to conventional type thickeners.

In Phase III, discussed in Sections 12-14, efforts were directed toward an eventual scale-up of the system. Sizing studies were done based on a flux plot developed to help estimate the settling area required for a desired level of final concentration. It was found that about 59% of the area required by conventional thickeners would achieve the same amount of separation and concentration with the use of the new dewatering system. Alternately, for about the same settling area, the degree of concentration achieved is about 60% greater.

Economic studies were also performed by the Engineering Staff of the Tennessee Valley Authority (TVA), at the request of EPA; these studies showed reductions in capital costs of about 10%, and reductions in annual operating costs of about 6%, for the entire FGD waste management system, not just the dewatering system.

To promote the technology transfer leading to eventual use of the system on a commercial scale, equipment design, layouts, etc., for a large scale evaluation of the dewatering system were done. In addition, studies were performed on the effects of shear on the settling behavior of the sludge. It was hypothesized and experimentally established that vigorous mixing of the FGD wastes causes a more rapid settling and a greater final concentration. The results of the experiments led to selection of a pump (to be used for the recycle stream from the clarifier to the thickener) that would vigorously mix the slurry causing it to exhibit more desirable settling characteristics.

In Phase IV, discussed in Sections 15-18, the work done during final shakedown testing and final pilot plant evaluation of the new dewatering system is discussed. A new concept in thickener operation (called "bang-bang" operation) that is based on the observations made during the pilot plant evaluation is outlined in detail in Section 17. The overall conclusions drawn from the pilot plant evaluation are listed in Section 18.

Following the shakedown test, several modifications were made to the pilot unit. The main modification made was the addition of more settling plates in the clarifier. The other modifications made had to do with operational considerations such as safety and ease of operation.

The initial goal of the pilot tests was to establish a baseline for comparison. At the baseline conditions, the underflow concentration was about 45 wt.% solids; the underflow rate was set at about 2.8 to 3.0 gpm. The average feed rate was about 13 to 14 gpm, with a solids concentration of about 15 wt.% solids.

Prior to start-up the thickener was filled with solids from the existing dewatering unit; the clarifier was not prefilled with solids. As a result, during the initial period of operation, solids accumulation took place in the clarifier.

As the clarifier began to fill with solids, it became necessary to increase the thickener underflow rate periodically to maintain a clear clarifier overflow. It was discovered that the feed rate to the system was being periodically increased and a corresponding increase in thickener underflow rate was necessary to keep the velocity in the clarifier below about 0.12 ft/min. By adding a splitter in the feed-stream, the feed rate was held fairly constant at about 13 to 14 gpm, and the thickener underflow rate could then be set as low as possible without plugging the underflow lines.

Each time the thickener underflow rate was decreased the underflow concentration increased significantly. When the thickener underflow rate was decreased from the baseline value of ~ 2.8 gpm to a value of less than 1 gpm, the underflow concentration increased to as high as 56 wt.% solids.

No significant change in the clarifier concentration was observed when its underflow rate was decreased from ~20 gpm to as low as 1.5 gpm. However, a significant response in separation efficiency was observed; it increased significantly each time the clarifier underflow rate was decreased.

While operating at a very low thickener underflow rate (~14 gpm) and a high underflow solids content (~53 wt.%), the underflow rate was increased significantly for about two hours. During this time, about 4 feet of concentrated sludge blanket was dumped. The underflow rate was then reset at its original low setting. This period of rapid dumping, followed by resetting the underflow rate to its original low level, had little effect on the underflow solids content. This was because of the relatively constant solids concentration in the large compression zone in the thickener.

On the basis of this observation and the observed strong inverse dependence of underflow concentration on steady-state (continuous) underflow rate, the concept of "bang-bang" thickener operation waas proposed. In this mode of operation, the thickener underflow rate is set as low as possible without the underflow lines plugging. Periodically, the underflow rate is stepped up for a short period in order to maintain an overall solids flowrate within the thickener. Using this mode of operation it should be possible to maintain a maximum underflow concentration.

The dewatering system was observed to be operationally versatile. Any possible plugging in the clarifier due to its overdesign, for example, should be easily avoidable by simply increasing the clarifier underflow rate. This could be done without having to lower the underflow concentration because further concentration would occur in the thickener.

Section 2

INTRODUCTION

Based on the 1973 consumption rate, the amuont of coal used for electric power generation in 1985 is expected to increase by 50%. Much of the abundunt coal resources of the nation, amounting to 60% of the total world supply, contain high amounts of sulfur. The burning of coal accounts for 50% of all sulfur dioxide emissions; these emissions can create a health hazard by significantly polluting the environment (Train, 1976). This growing environmental problem has stimulated a large amount of research and development on methods for removing the pollutant from waste gases. Among these methods, the lime/limestone processes have the outstanding advantages of relative simplicity, low capital investment, and freedom from the problems of marketing a by-product. This non-regenerable method of flue gas desulfurization (FGD) has been favored by the electric power industry as the most economical method available. In this method, sulfur dioxide in the flue gases is removed by scrubbing with a lime/limestone slurry. Large amounts of FGD by-product waste are produced.

To aid in the disposal of the FGD waste (a major problem with FGD wastes is physical stability for handling and disposal, especially in landfilling), the sludge should be concentrated and the clarified water recycled as much as possible. The waste from the scrubber is fed to a settling unit, which separates it into two streams: one containing large amounts of solids, and the other free of solids. Clear liquid is recycled to the scrubber system. The thick FGD waste is ponded or used as a landfill after necessary intermediary treatment (Goodwin, 1978).

The effectiveness of the dewatering step depends on the nature of the FGC waste which includes such factors as concentration of solids, particle size range, density of solids, the chemical nature of solids, density and viscosity of the fluid, etc. and the design of the dewatering equipment.

The original objective and approach for this project were stated as follows:

> The objective is to develop more economically feasible methods than those currently available for continuous treatment of flue gas cleaning (FGC) wastes. Initial efforts are to be concentrated on applying gravity sedimentation to rapidly, inexpensively dewater FGC wastes. The approach will be to: Develop methods for characterizing the physical properties and behavior of FGC wastes. Develop procedures based on the models presented herein to allow accurate design and analysis of gravity sedimentation units used in treating FGC wastes. Perform bench scale and pilot tests to evaluate the feasibility of using gravity sedimentation units for treating FGC wastes. Perform developmental studies to determine the best design for channel promoters. Bench scale and continuous pilot tests are to be performed to evaluate the feasibility of such devices for improving dewatering efficiency while treating FGC wastes. Assess filterability of different types of FGC sludges dewatered to different degrees and pumping requirements for the different sludges.

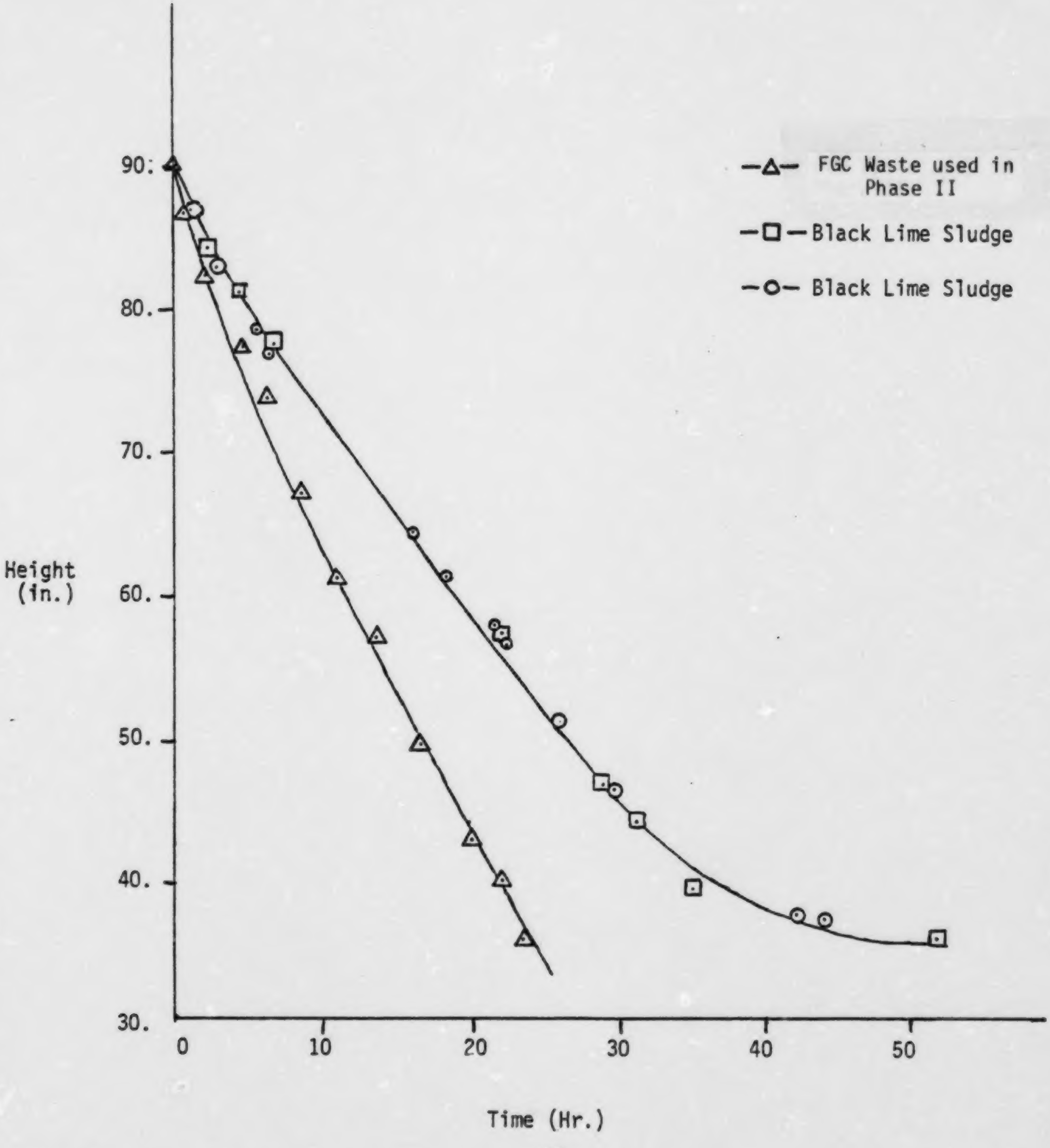
Section 3

PHASE I

PRELIMINARY STUDIES

Initially, batch settling and filtration studies were performed to evaluate the physical nature of the FGC waste. The FGC waste used in these studies was obtained from the Louisville Gas and Electric Company (a high sulfite content byproduct of the carbide lime process for making acetylene). The sludge is characterized by the settling curve given in Figure 3.1, in which a comparison is made with the settling behavior of the FGC waste used in Phase II studies. A significant concentration profile was observed to exist across the sludge blanket during batch settling; also, the path of the solid/liquid interface during settling (the batch settling curve) was found to be of the classical shape that is indicative of hindered settling. These observations implied that FGC waste is highly compressible. This implication was confirmed by the results of the filtration studies, in that the specific cake resistance varied significantly when the pressure drop through the cake was increased. In addition, the concentration of the waste sludge existing at the bottom of a settled blanket of the sludge was observed to depend on total sludge blanket depth: the deeper the sludge blanket, the higher the concentration of solids in the bottom layer. Thus, the compressive resistance of the waste sludge was concluded to be sufficiently low so that a layer of sludge in a sludge blanket can be compressed by the weight of the sludge overburden alone. Based on these observations then, the depth of the sludge blanket in a continuous settler should have a significant effect on dewatering efficiency. This is shown to indeed be true in Phase II

Figure 3.1. Batch Settling Curves for FGC Waste Sludges



results. For convenience of reporting, all detailed settling data are reported in the Phase II discussion. The general results discussed above were sufficient to justify proceeding with the gravitational sedimentation work reported in the Phase II discussion.

A design study was conducted, verifying the utility of Kynch's theory for designing thickeners. This approach, however, is severely limited in that it is not accurate when compressive settling is dominant. This work was published in the "Fundamental Aspects of Hydrometallurgical Processes Section" of the 1978 AIChE Symposium Series, Vol. 74, No. 173. Refer to Appendix A for a copy of this manuscript.

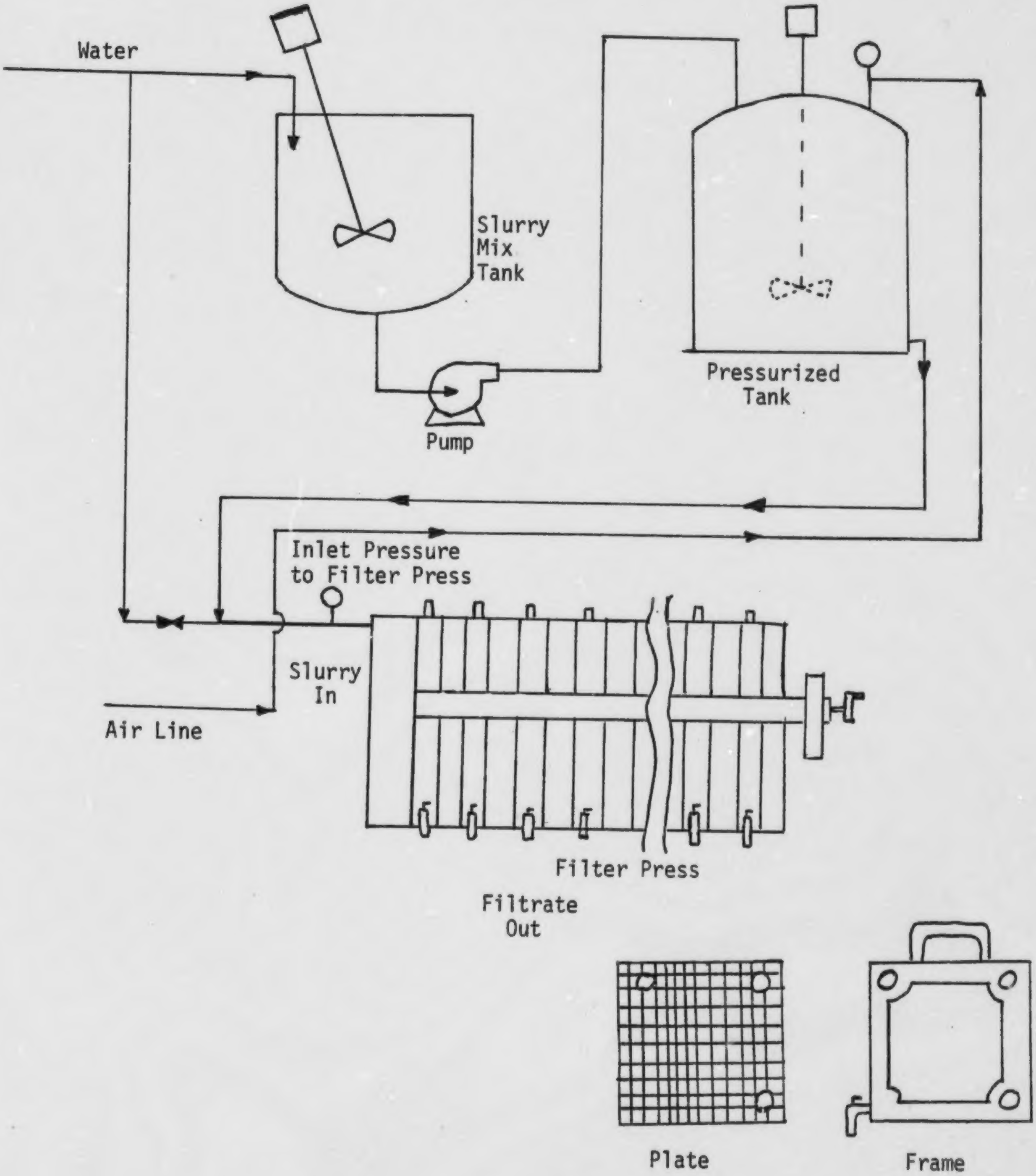
The filtration unit used is diagramed in Figure 3.2. The filter is a one square foot plate and frame filter. Sample calculations, showing typical data work up, for a filtration run are given in Appendix B. V_m is the filter cake volume necessary to produce all additional pressure drops (besides that across the cake) between the points across which the total pressure drop (ΔP) is measured. C_v is defined as follows:

$$C_v = \frac{K_p S_v^2 (1-\varepsilon)^2 \mu \rho w}{2 g_c \varepsilon^3 [\rho_s (1-\varepsilon)(1-w) - \varepsilon \rho w]}$$

where,

kp ≡ proportionality constant
S_v ≡ surface area/volume of particles
ε ≡ cake porosity
μ ≡ filtrate viscosity
ρ ≡ filtrate density
W_s ≡ weight fraction of solids
g_c ≡ gravitational constant
ρ_s ≡ density of solids

Figure 3.2. Filter Press Apparatus


Water
Slurry
Mix
Tank
Pressurized
Tank
Pump
Inlet Pressure
to Filter Press
Slurry
In
Air Line
Filter Press
Filtrate
Out
Plate
Frame

The following two conclusions can be made from the data given in Tables 3.1 and 3.2:

- The cake resistivity, α, is not a strong function of slurry concentration at the higher ΔP values.
- The cake is highly compressible; C_v and V_m both vary with changes in ΔP.

Based on these preliminary results, Phase II gravity sedimentation studies were started. The overall strategy was to increase the dewatering efficiency of the gravity sedimentation units in the dewatering train. This would result, in turn, in a decrease in the dewatering load for the filters.

Table 3.1. Preliminary Filtration Data*

Cake Porosity ε	Pressure Drop ΔP (psi)	Wt. Fraction Solids in Feed W (10^3)	Filtration Functions		Cake Resistivity α
			C_v (10^{-6} $\frac{lb_f\ sec}{ft^4}$)	V_m (10^3 ft^3)	(10^{-14} $\frac{ft}{lb}$)
0.75	7	4.2	7.13	2.96	9.9
0.84	7	7.9	9.14	1.98	15.5
0.77	7	15.8	20.9	2.13	4.6
0.78	7	39.8	36.9	1.13	0.71
0.71	15	4.3	6.58	2.48	24.4
0.73	15	6.6	13.3	3.59	15.5
0.75	15	2.1	24.0	1.90	3.2
0.71	15	2.8	26.5	1.46	2.6
0.72	22	7.8	14.7	3.24	7.4
0.77	22	17.8	32.0	3.02	8.1
0.71	30	8.8	20.0	3.58	3.5
0.78	30	26.0	65.7	3.94	3.9

* Note: Computation procedures for filtration functions (C_v,V_m) and definitions are given in Appendix B.

Table 3.2. Preliminary Filtration Data*

Cake Porosity ε	Pressure Drop ΔP (psi)	Wt. Fraction Solids in Feed W (10^3)	Filtration Functions	
			C_v (10^{-6} $\frac{lbf\ sec}{ft^4}$)	V_m (10^3 ft^3)
0.74	20	7.9	3.6	0.28
0.79	20	8.1	4.0	0.28
0.84	7	7.8	1.14	0.44
0.71	15	4.3	0.82	0.69
0.72	22	4.6	1.04	0.48
0.75	15	21.	3.0	0.09
0.77	15	28.	3.3	0.08
0.78	22	18.	4.0	0.23
0.71	30	26.	8.2	0.23
0.75	7	4.2	0.89	0.28
0.78	7	40.	4.6	0.02
0.73	20	22.8	10.4	0.04
0.74	20	12.1	5.1	0.11
0.52	30	15.1	8.0	0.04
0.72	30	15.1	8.4	0.05
0.66	30	6.7	2.8	0.04

* Note: Computation procedures for filtration functions (C_v,V_m) and definitions are given in Appendix B.

Section 4

PHASE II

NEW GRAVITY FGD WASTE DEWATERING SYSTEM: SYNTHESIS AND DEVELOPMENT

INTRODUCTION

Thickening has long held the status of an engineering science but is still a long way from being an exact science; there is considerable 'art' involved in deciding certain points (Roberts, 1949). In 1916, Coe and Clevenger, using batch settling tests, first developed a method to calculate needed thickener areas and started to deal with thickening by scientific approach. However, the design of dewatering equipment still has not changed substantially over a long period of time. Conventional sedimentation units, consisting of a large concrete or steel tank, 200 to 300 feet in diameter, and a rake mechanism to aid the discharge of thick underflow, have been widely used for dewatering up to now.

Many investigators since Coe and Clevenger, such as Richardson and Zaki (1954), Peter Kos (1974, 1975), Shirato (1970), Michaels and Bolger (1962), etc., have found a lot of useful results concerning the nature of thickening sludges. Consequently, the thickening mechanism is now better understood.

For the flue gas desulfurization process, the nature of the sludge depends on the type of scrubbing operation. The sedimentation characteristics and landfill suitability of the sludge are improved by more complete oxidation of lime/limestone to $CaSo_4 \cdot 2H_2O$ (gypsum) rather than $CaSO_3 \cdot 1/2H_2O$ (Goodwin, 1978). The capital and operating costs for the filtration operation are very high. Minimization of the filter loads or complete elimination of the filtration operation is desirable.

Two primary approaches have been taken to improve the dewatering efficiency of the clarifier, thus minimizing FGD waste treatment costs. One is the chemical approach, for example, as discussed above, which gets more complete oxidation of the FGD waste to gypsum rather than $CaSO_3 \cdot 1/2H_2O$. The other is the design approach, which is the subject of the present work. Two classical types of gravity settlers were studied and a novel conceptual design was developed by combining a clarifier with a thickener. The combined dewatering system can concentrate FGD waste sludge to 40 to 60 weight per cent of underflow concentration while yielding a clear overflow.

The following two classical types of gravity settlers were studied:

1. The laminar-type clarifier, in which inclined plates or tubes are used to increase the settling surface area and the settling volume is divided into many shallow chambers;
2. The circular basin type of thickener, in which the depth of the concentrated waste can be varied depending on operational conditions.

In the first type of settler, free settling occurs, while in the second type compressive settling can be achieved by maintaining a sufficient depth of the sludge (Goodwin, 1978; Fitch, 1975 -part 1,2, and 3). Compressive behavior results from the solids waste overburden on the underlying layers in the basin. The dewatering efficiency of the laminar-type of clarifier depends on the angle of inclination of the tubes or plates, the location of the feed port, the feed concentration, and the feed rate. The effect of these parameters on the clarification efficiency and on clarifier underflow concentration were studied. For the standard circular-basin type of thickener, dewatering efficiency depends on feed port location, depth of the concentrated wastes accumulated on the bottom, underflow rate, and solids loading. The effect of these parameters and the importance of compressive studies performed for both the laminar-type thickener and the circular basin type of thickener helped to determine the function and usefulness of each unit, keeping in view the overall treatment objectives: concentrating the wastes as much as possible without using any chemical additives or expensive treatment method, like filtration or centrifugation, while maintaining a clear overflow.

Although the laminar-type clarifier can be used to improve separation efficiency significantly by itself, solids loading is largely controlled by the area provided in either the clarification zone or the critical settling zone (Fitch, 1975-part 1). According to Fitch's derivation, solids handling capacity ($\dot{F}$) has a critical value ($\dot{F}_c$) for a certain area of thickener. If the solids loading is above this value, i.e., $\dot{F}_f > \dot{F}_c$, then the solids cannot pass through the critical zone. A zone of constant concentration builds up and propagates from the bottom to the top of the settler. The underflow draw-off rate and concentration thus become limited by the critical solids-handling capacity ($\dot{F}_c$). If provision is made for additional area by the combining of the laminar-type of clarifier with the circular-basin type of thickener and recycling the underflow from the clarifier into the thickener, the critical solids-handling capacity will be switched to a higher value, and the underflow concentration and the overall separation efficiency will be improved. Moreover, such a system has a greater degree of controllability than the conventional dewatering systems using a combined laminar-type clarifier and a thickener without recycling.

The overall dewatering system presented here consists of both a clarifier and a thickener with recycle of the underflow from the clarifier back into the thickener. Different system configurations were considered, and simple steady-state analyses were used to evaluate the effect of the relative loading of the two parts, which is under the control of the designer and the operator. The wide variations in feed rates and concentrations, the design limitations, and the chemical and physical property variations encountered in practice imply that fine tuning during operation is very essential to achieving optimum operation and desirable results.

Section 5

LITERATURE REVIEW

In nearly all commercial cases, gravity settling operates continuously. The suspension is fed into one or more settling basins or chambers, and while it is passing through, solids settle out. The solids are removed together with a fraction of the liquid, as thickened underflow. The rest of the liquid overflows, more or less clarified or depleted of solids. Equipment structures used for this purpose are commonly called thickeners or clarifiers.

In order to approach settling in a scientific manner, one should first know the settling behavior. Usually, particles may settle in four markedly different manners; which one they follow is governed by their concentration and by the relative tendency of the particles to cohere. These different types are represented paragenetically in Figure 5.1 (Fitch, 1977). At low concentrations the type of sedimentation that exists is called clarification. Particles are, on the average, far apart and free to settle individually; but collisions occur. In this regime there are two different types of settling: particulate and flocculating, depending on whether the particles cohere or not. As particles become more concentrated they exhibit a zone settling regime; they are in contact with neighbors and tend to link into a plastic structure. At some higher concentration the pulp structure becomes so firm that it develops compressive strength. Each layer of pulp solids is able to provide mechanical support for layers above, and each layer is subjected to a solids stress or squeeze. When this happens the pulp is said to be in compression.

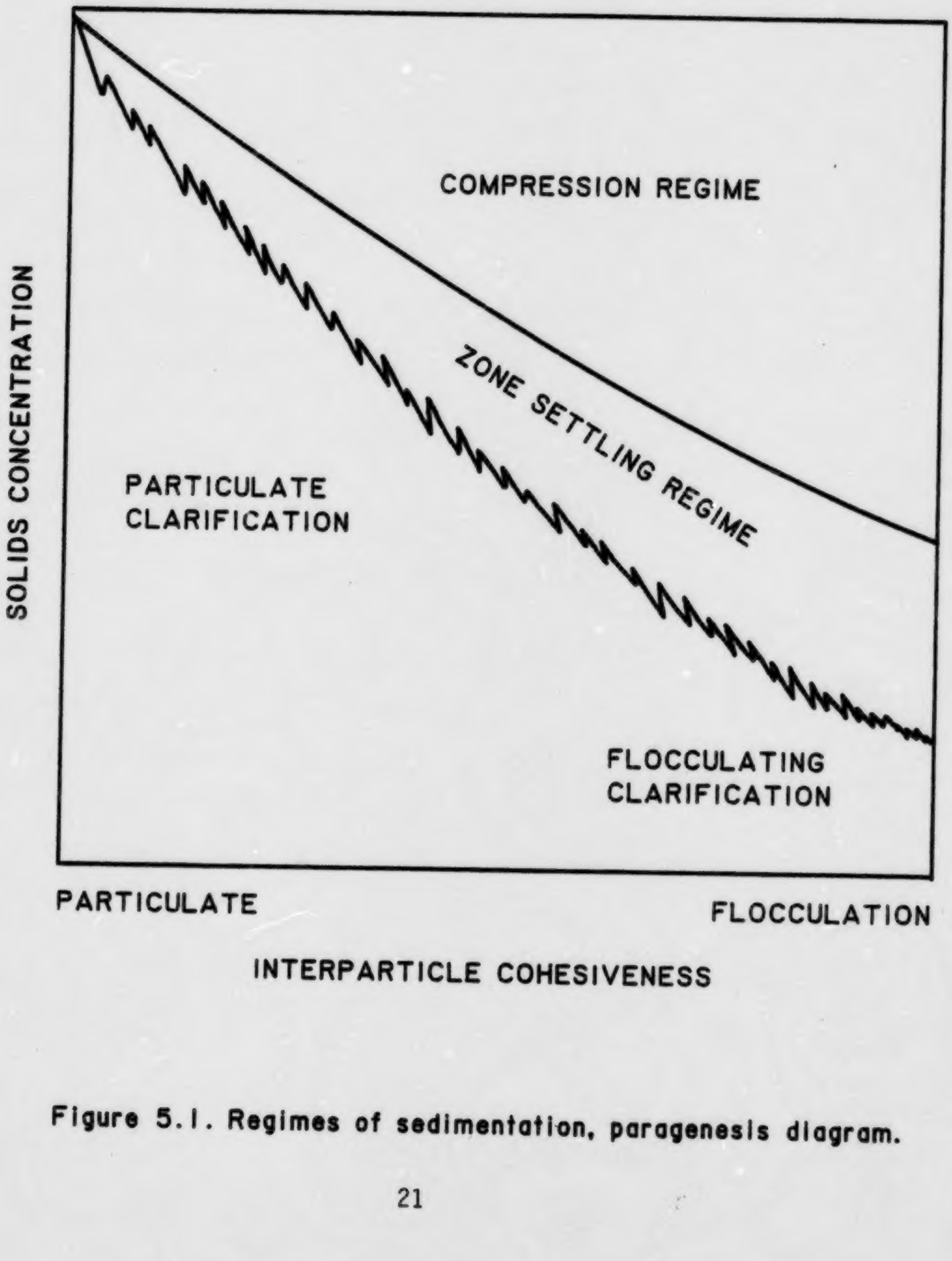

COMPRESSION REGIME
ZONE SETTLING REGIME
PARTICULATE CLARIFICATION
FLOCCULATING CLARIFICATION
SOLIDS CONCENTRATION
PARTICULATE
FLOCCULATION
INTERPARTICLE COHESIVENESS

Figure 5.1. Regimes of sedimentation, paragenesis diagram.

Thickening

This comprises critical zone settling and compression. In 1961, Coe and Clevenger first showed the existence of concentration zones within a thickener. By comparing the settling process occurring in both batch settling and continuous thickening, they formulated a basic equation for sizing thickeners from batch settling results. If the solids flux $\dot{F}$ is given by:

$$\dot{F} = C(U + V_u) \tag{1}$$

where

$c \equiv$ solids concentration

$u \equiv$ solids settling rate at a solids concentration of c

and $V_u \equiv$ bulk velocity resulting from underflow withdrawal

by material balance, noting that $V_u = Q_u/A$

$$\dot{F} = C(U + Q_u/A) = C_uQ_u/A \tag{2}$$

and eliminating Q_u/A, then

$$\dot{F} = \frac{U}{(1/C) - (1/C_u)} \tag{3}$$

1. Zone Settling

According to Coe and Clevenger's procedure, settling rates can be determined for various C's in batch tests, and the results plugged into the above equation. Their procedure will give, for any fixed value of C_u, a set of thickener fluxes ($\dot{F}$) corresponding to the different concentrations (C). If the results are graphed, they will appear as in Figure 5.2. There

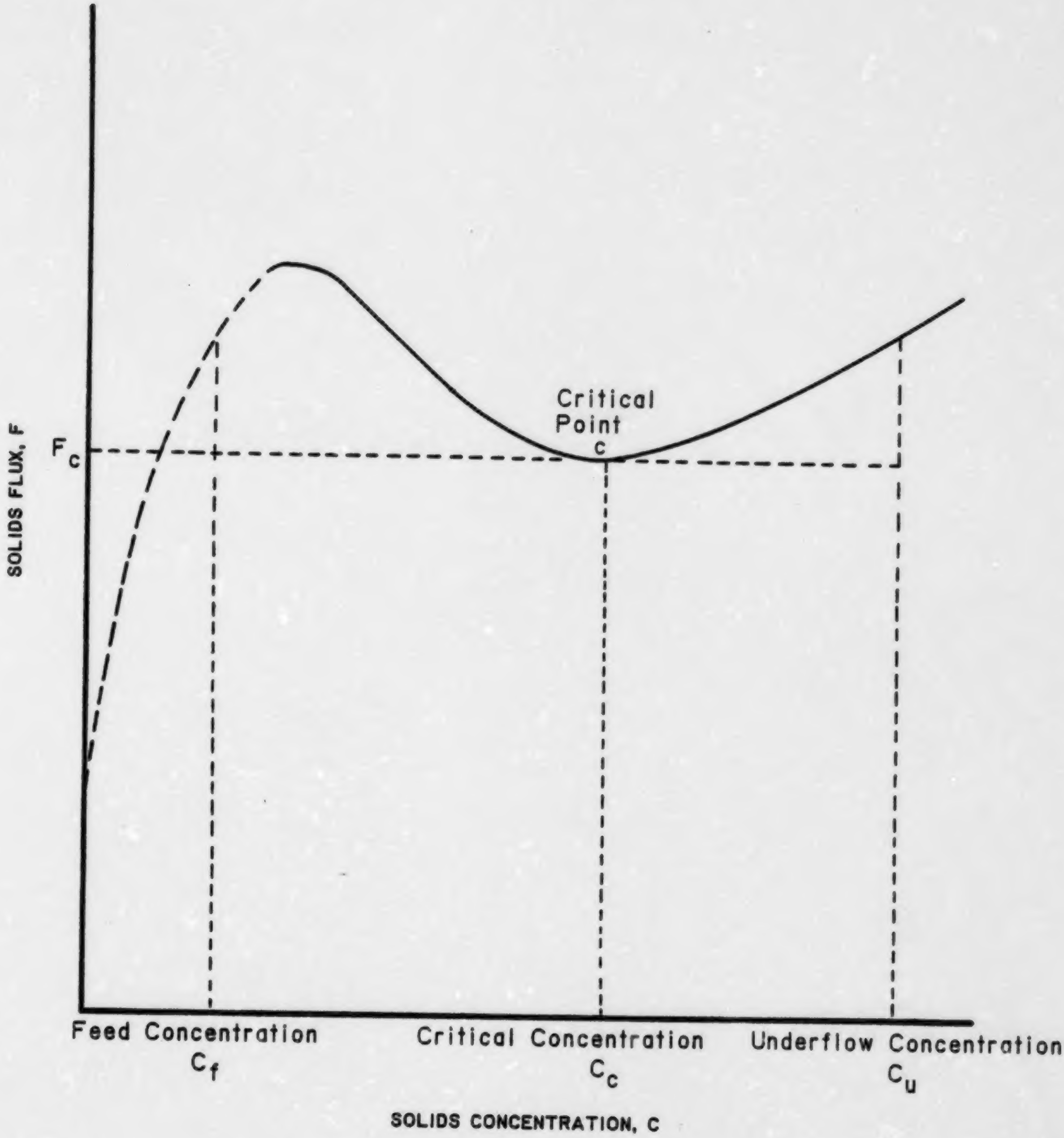


Figure 5.2. Solids flux vs concentration.

is a minimum $\dot{F}_c$ at some concentration C_c. It is normally assumed that this is the maximum flux that can pass to the underflow. At solids loadings greater than this, a critical zone of concentration (C_c) will form in the thickener. The solids that cannot pass this zone, as fast as they are supplied, add to the zone, increasing its height until it fills the thickener and the excess solids overflow. $\dot{F}_c$, therefore, constitutes an upper boundary for solids loading. Thus zone settling imposes an area demand on thickener design.

On the other hand, when feed solids flux drops below the critical level, solids pass down through the zone faster than they are being replenished at the top. The zone dwindles and disappears. Thus, no particular depth is necessary to accomodate zone settling. It imposes no depth demand on design.

According to COE and Clevenger's batch settling test, a column of wastes, initially of some uniform concentration in the thickening regime, is allowed to settle. The level H of the interface is measured as a function of time, as shown in Figure 5.3. The two slope discontinuities divide the subsidence curve into a constant-rate section from a to b, a first falling-rate or transition section, b to c, and a second falling rate section beyond c. The constant-rate section corresponds to wastes settling at their initial concentration. This provides the data used in the classical Coe and Clevenger procedure for determining flux curves and critical fluxes. During the final falling-rate section, the wastes are in compression.

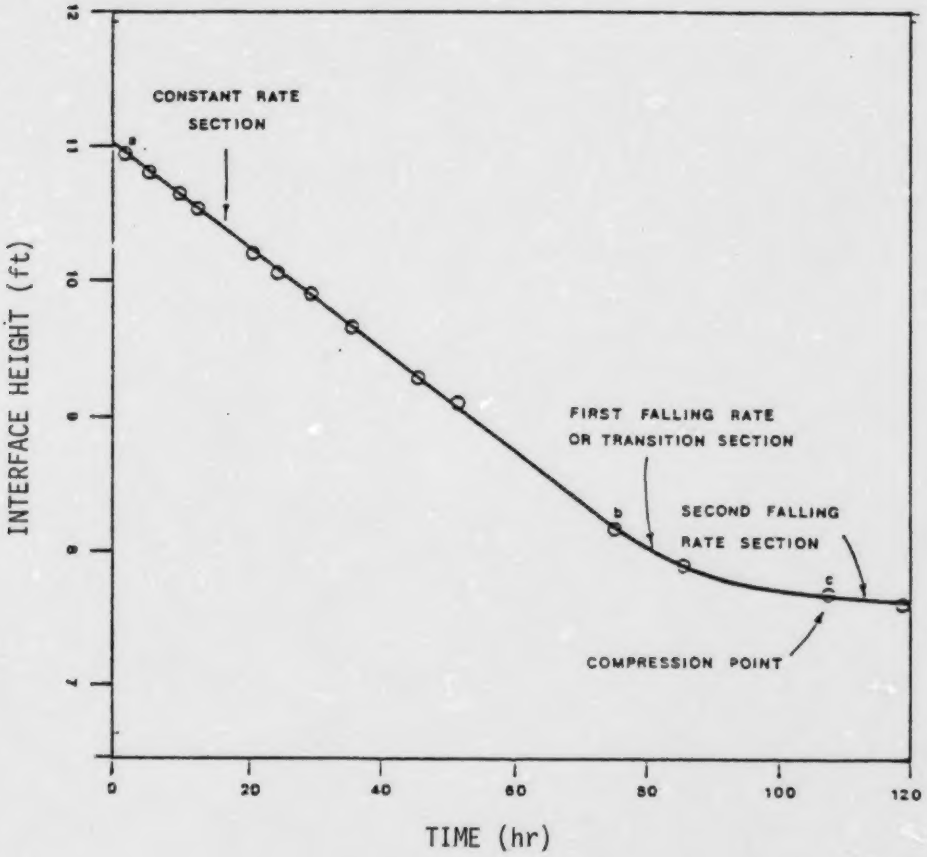


Figure 5.3. The batch settling test for FGD wastes, C_0 = 29.85 wt %.

2. Compression

In compression, settling rate (U) is no longer a function of concentration alone, but also of concentration gradient ($\partial C/\partial H$), and possibly of other variables.

Coe and Clevenger observed empirically that for certain metallurgical pulps, the underflow concentration attained in a compression zone was approximately a function of the residence time (θ_s). That is, $C_u = f(\theta_s)$. Therefore, compression imposes a solids detention demand or a demand on the volume of a compression zone.

Clarification

It was first reported by Boycott (1920) that solids suspensions in liquids settle faster in inclined test tubes than in vertical tubes under the same conditions. He also observed that initial settling rates increased with decrease of angle of inclination (to the horizontal) of the test tubes. In 1937 Nakamura and Kuroda first presented a quantitative theoretical model. They considered settling between two parallel inclined planes (Figure 5.4):

$$-\left(\frac{dH/dt}{U_v}\right)_o = 1 + \frac{H_o \cot\theta}{B} \tag{4}$$

where

B ≡ horizontal distance between planes
H ≡ height of suspension-liquid from the bottom
H_o ≡ initial height of suspension-liquid from the bottom
U_v ≡ vertical settling rate of suspension (in the absence of an inclined plane).
t ≡ time
θ ≡ angle of inclination

In the above equation the subscript o refers to initial conditions.

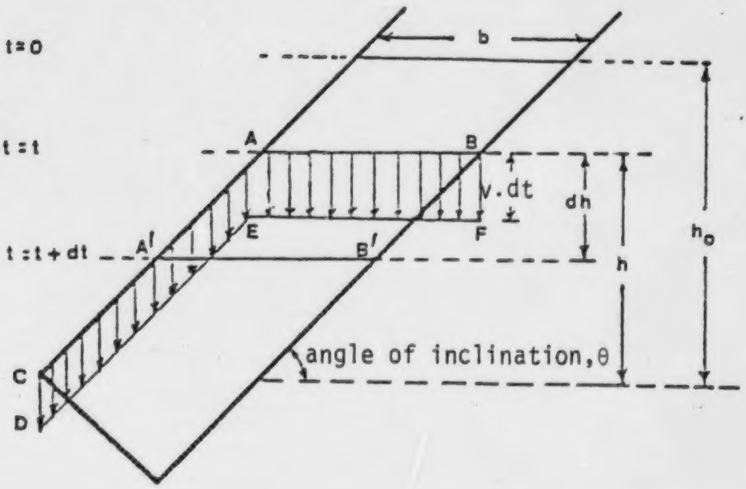

t=0
t=t
t=t+dt
b
A
B
A'
B'
E
F
C
D
v.dt
dh
h
h_o
angle of inclination, θ

Figure 5.4 Theoretical model of Nakamura and Kuroda.

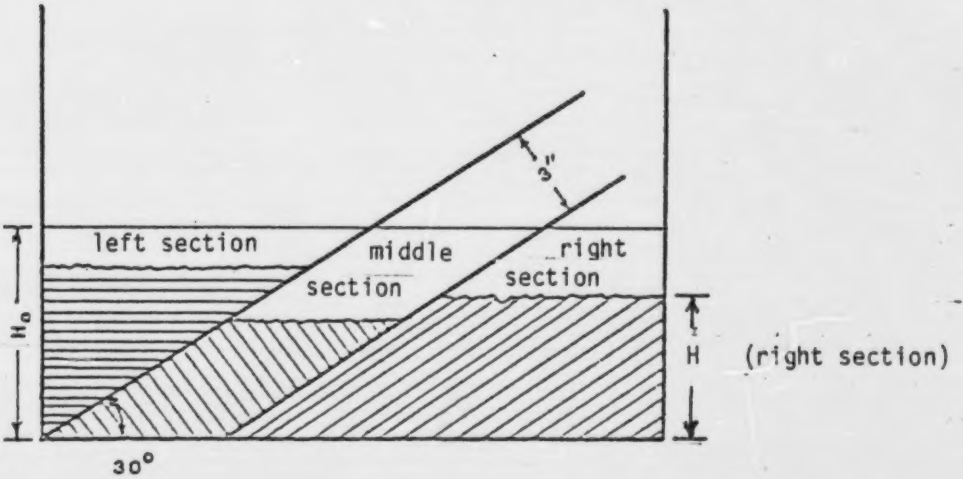

3"
left section
middle section
right section
H_o
H
(right section)
30°

Figure 5.5. Batch settling box with inclined plates, showing settled waste heights

They conducted settling experiments with blood suspensions in 61-mm-long square test tubes inclined at 45°. Fairly good agreement was obtained between the experimental results and their model. Following Nakamura and Kuroda's work, Eli Zliezer Rubin (1975) conducted an experimental and theoretical study on the effect of inclined planes on batch settling of solid-liquid suspensions by using vertical, right-angled trapezoid and parall[illegible]gram shaped vessels. They found that initial settling rates in the presence of inclined planes can be higher by a factor of up to 3 compared to settling rates in regular vertical vessels under the same conditions. The settling rates also increase with decrease of angle of inclination or distance between planes and with increase of initial suspensions height or suspension concentration. The typical identifications of the above are shown in Figures 5.5, 5.6, and 5.7, obtained by pouring slurry, which was waste material obtained from the Louisville Gas and Electric lime scrubber process, into a rectangular box in which two parallel plates are inserted at 30°.

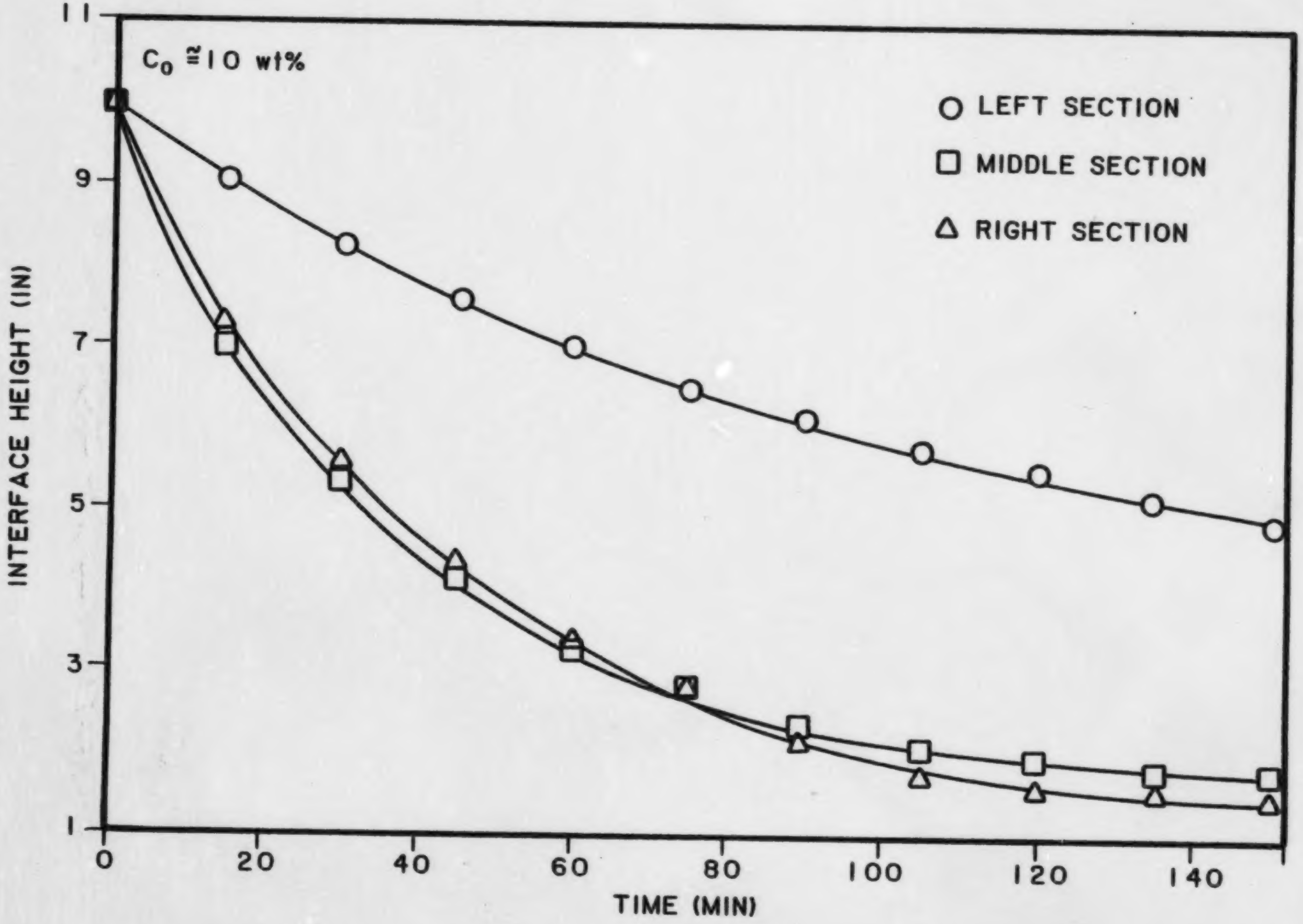


Figure 5.6. Batch settling tests in a settling box for FGD wastes.

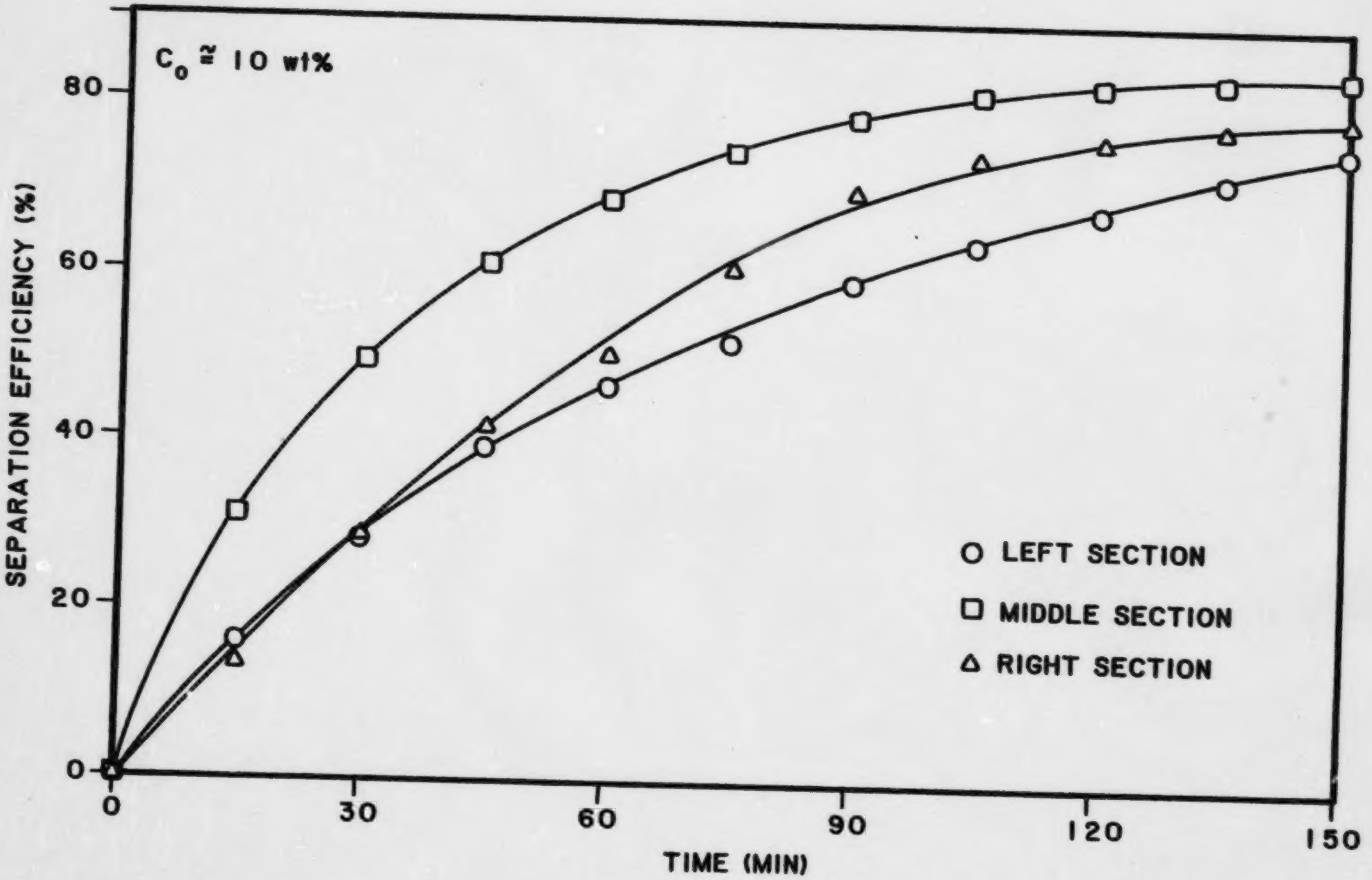


Figure 5.7. The separation efficiencies of the batch settling tests in a settling box.

Section 6

MATERIALS AND MEASUREMENT

The FGD waste used in all experiments in Phase II was obtained from the Louisville Gas and Electric Company, Louisville, Kentucky in April 1977. This waste was produced by the comapany's Unit No. 6, which uses a marble bed scrubber. The carbide lime used as the absorbing medium contained mostly CaO, and less than 1.5% of it was $CaCO_3$. The solids in the FGD waste contained almost 85% $CaSo_3 \cdot 1/2\ H_2O$, whereas about 15% was $CaSo_4 \cdot 2H_2O$ or $CaSO_4 \cdot 1/2\ H_2O$. The specific gravity of the solids was about 2.3.

Flow rates of the underflow and overflow streams, Q_u and Q_o, were measured on a volume-per-unit time basis by collecting the slurry in a graduated flask for a fixed length of time (10 minutes). In steady state operation a total slurry volume balance over the system is expressed by:

$$Q_f = Q_u + Q_o, \tag{5}$$

where

$Q_f \equiv$ Feed flow rate

$Q_u \equiv$ flow rate of the underflow

and $Q_o \equiv$ flow rate of the overflow

Whereas a volume of solids balance is

$$Q_f W_f = Q_o W_o + Q_u W_u. \tag{6}$$

where W_f, W_o and W_u are weight percent solids in the feed, overflow and underflow respectively.

For the determination of whether a system is in a steady-state condition, the following mass balance equations are used:

$$\dot{M}_{in} = (Q_o + Q_u)\cdot C_f \tag{7}$$

$$\dot{M}_{out} = (Q_o + Q_u)\cdot C_f \tag{8}$$

where, $\dot{M}_{in} \equiv$ Mass flow rate in

$\dot{M}_{out} \equiv$ Mass flow rate out

and $$C = \frac{W\cdot\rho_s}{W + \frac{\rho_s}{\rho_\ell}(1-W)} \tag{9}$$

where W = wt. % solids

ρ_s = solids density

and ρ_ℓ = liquid density

$\therefore$ Percentage approach to steady state

$$= \dot{M}_{out} \times 100\%, \text{ if } \dot{M}_{out} < \dot{M}_{in} \tag{10}$$

$$= \left(1 - \frac{\dot{M}_{out} - \dot{M}_{in}}{\dot{M}_{in}}\right) \times 100\%, \text{ if } \dot{M}_{out} > \dot{M}_{in} \tag{11}$$

Solids flux $(\dot{F})$

$$= \frac{Q\cdot C}{A} \tag{12}$$

For the present work, $\rho_\ell = 1.0$ gm/ml, $\rho_s = 2.3$ gm/ml.

If the overflow is clear, W_o is negligible in the above equations.

All the measured separation efficiencies are defined as the volume of overflow separated in a certain time divided by the volume of feed; i.e.,

separation efficiency (%)

$$= \frac{Q_o}{Q_f} \times 100\% \qquad (13)$$

The concentrations of all streams were obtained by drying a small amount of sample to drive out the free water in the slurry and reweighing. The weight percent solids was obtained by finding the ratio of the weight of solids to the weight of slurry. Care was taken to see that the water of hydration was not lost during drying. The total time of drying under a hot lamp was maintained at 10 to 15 minutes for about 10 ml of sample. The results obtained by this drying method were compared with those obtained by vacuum drying. The comparison was made by using two samples of the same volume from the same slurry and by observing the change in weight of the sample with time. As shown in Figure 6.1, the weight per cent of the solids as calculated from the two drying methods approaches the same value.

The flow patterns of the slurry were checked in all the experiments by dye tests. The details of the dye tests are available elsewhere (Kleinschrodt, 1975).

In batch settling experiments, the solids concentration profile along the height of the column was measured by use of a nuclear density gauge of Ohmart. The calibration and measurement details were developed earlier (Kennedy, 1976).

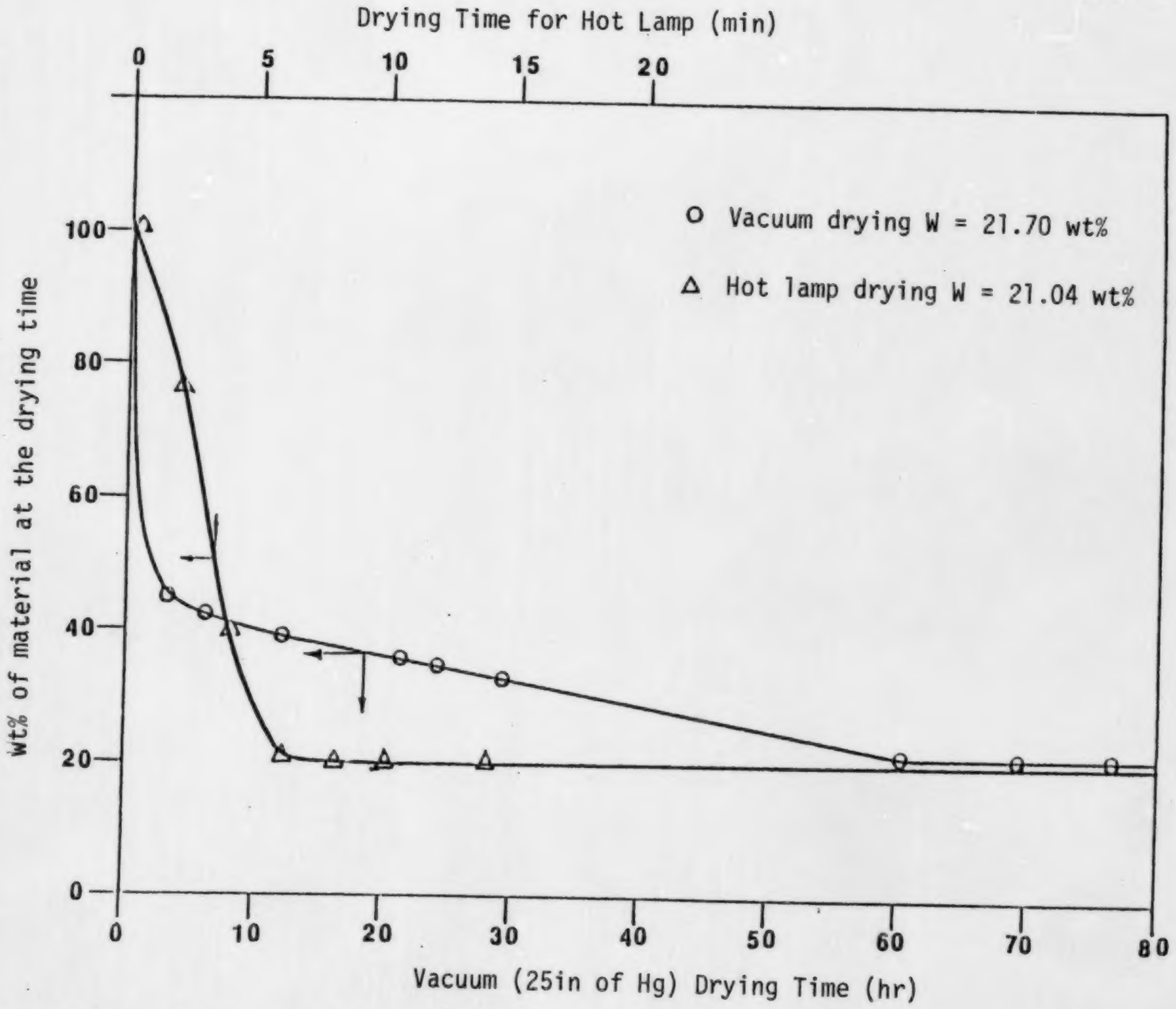


Figure 6.1. Comparison between hot-lamp drying and vacuum drying for determination of solids concentration.

Section 7

DESIGN AND OPERATIONAL LIMITATION: CLARIFICATION CONTROLLED OR THICKENING CONTROLLED

Classically the initial step in the design of dewatering systems is to determine whether clarification or thickening is the limiting consideration. The dewatering unit - whether it be a clarifier or thickener - usually has two functional roles: clarification and thickening. In the present work, however, these two functional roles are treated separately, or decoupled as much as possible, because the laminar-type clarifier, having more area than the circular-basin type of thickener, can separate faster in the free-settling regime, whereas the latter has a greater compressive force to squeeze the water out of the sludge. Therefore, the total separation efficiency of the system using a combined laminar-type clarifier and a thickener should be greater when the two functions are operated separately than when operated together in a clarifier and a thickener. Nevertheless, settling area is normally the main consideration in the design of a dewatering unit. Whether more area is required for clarification of the overflow or for thickening to the desired underflow concentration is determined by the designer.

There are several different methods, for example, that can be used to estimate the area necessary for clarification requirements: the long tube test, the short tube test, and second-order-bulk settling (Fitch, 1977). Clarification requirements, rather than thickening requirements, have been observed during continuous settling studies to be limiting only in that the desired clarity, which could always be attained by increasing the settling area, is limited by the clarifier's performance and efficiency. As a result, no specific tests for determining clarification requirements, such as those mentioned above, have been performed. The classical design approach would thus be to determine the area requirements for thickening, and, more or less, ignore the clarification role that the clarifier would perform. The reason is that if there is sufficient area for attaining the desired degree of thickening, clarification requirements would also be met. Therefore, a clarifier is used in this work, even though clarification is not limiting. Its role, however, is not only to fulfill clarification requirements, but, more importantly, to enhance the performance of the thickener, so that the FGD wastes can be thickened to very high concentrations of solids with less settling area than that required by the use of classical types of thickeners.

The method standardly used for estimating area requirements for a classical type of thickener involves the use of flux plots. Such a plot for an FGD waste (Figure 7.3) was obtained from the Louisville Gas and Electric Company (April, 1977). For the construction of this plot, several batch settling tests were performed. For these batch settling tests, a 1' I.D., 12' long column was used: this size of column is more than adequate to minimize secondary interferences (e.g., wall effects). In each batch settling test, a column of waste, initially of some uniform concentration, obtained by stirring with air bubbles, was allowed to settle, and the level H of the solid-wastes interface was measured as a function of time. Five representative tests for a long column (11 ft.) are shown in Figure 7.1, and two for a short column (1 ft.) are shown in Figure 7.2. Thus, the initial settling rates ($-(dH/dt)_0$) are calculated from these curves. The results are shown in Table 7.1 and Figure 7.3. Figure 7.3 is discussed in further detail later in this section. Some dispute exists as to whether the settling rate in the constant rate section or that in the first falling rate (or transition) section, as indicated in Figure 5.7, should be used in the construction of the flux plot.

For over 50 years, the Coe and Clevenger method of batch settling (for which several tests are needed) has been used to calculate needed thickener area. But Kynch modified this method and made it possible for one to find the required solids flux plot from one single batch test. In fact, these two methods should give the same solids flux plots if Kynch's assumption is valid (the settling rate, U, is solely a function of the local solids concentration). That is, the slurry exhibits ideal properties. However, most real slurries exhibit non-ideal properties; a number of researchers have shown that additional factors besides solids concentration affect the settling rate (Shannon and Tory, 1965; Scott, 1968-a,b,1970). Therefore, the settling rates measured from the constant rate section (Coe and Clevenger procedure) are different from those obtained from the first falling rate section (Kynch procedure) for any specific concentration. In general, the use of settling rates measured in the constant rate section leads to underestimation of the thickener area, whereas those measured in the first falling rate section lead to overestimation of thickener area. As a conservative measure in estimating the possible reduction in thickener

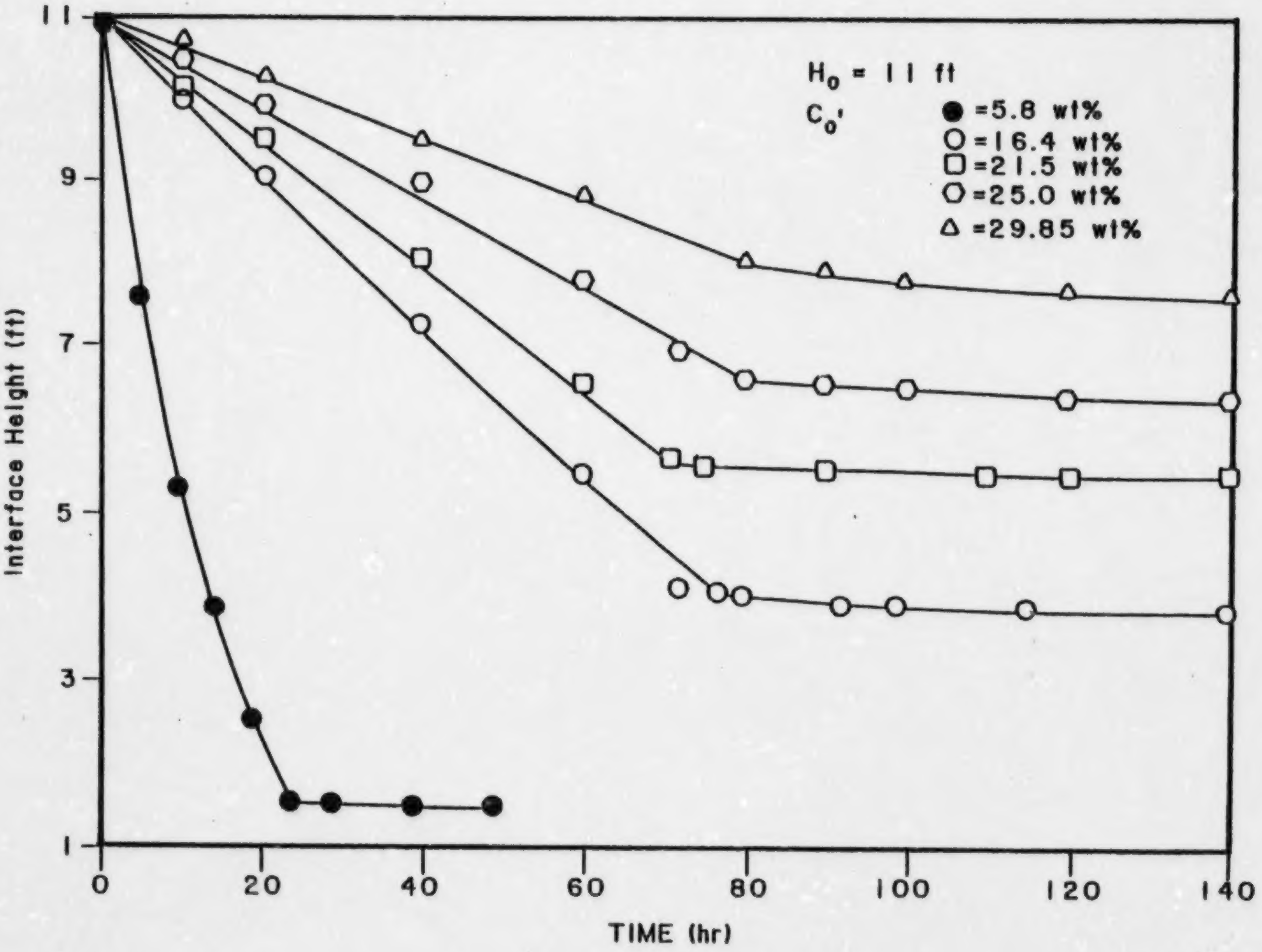


Figure 7.1. Batch settling curves obtained in a long column (H_0 = 11 ft).

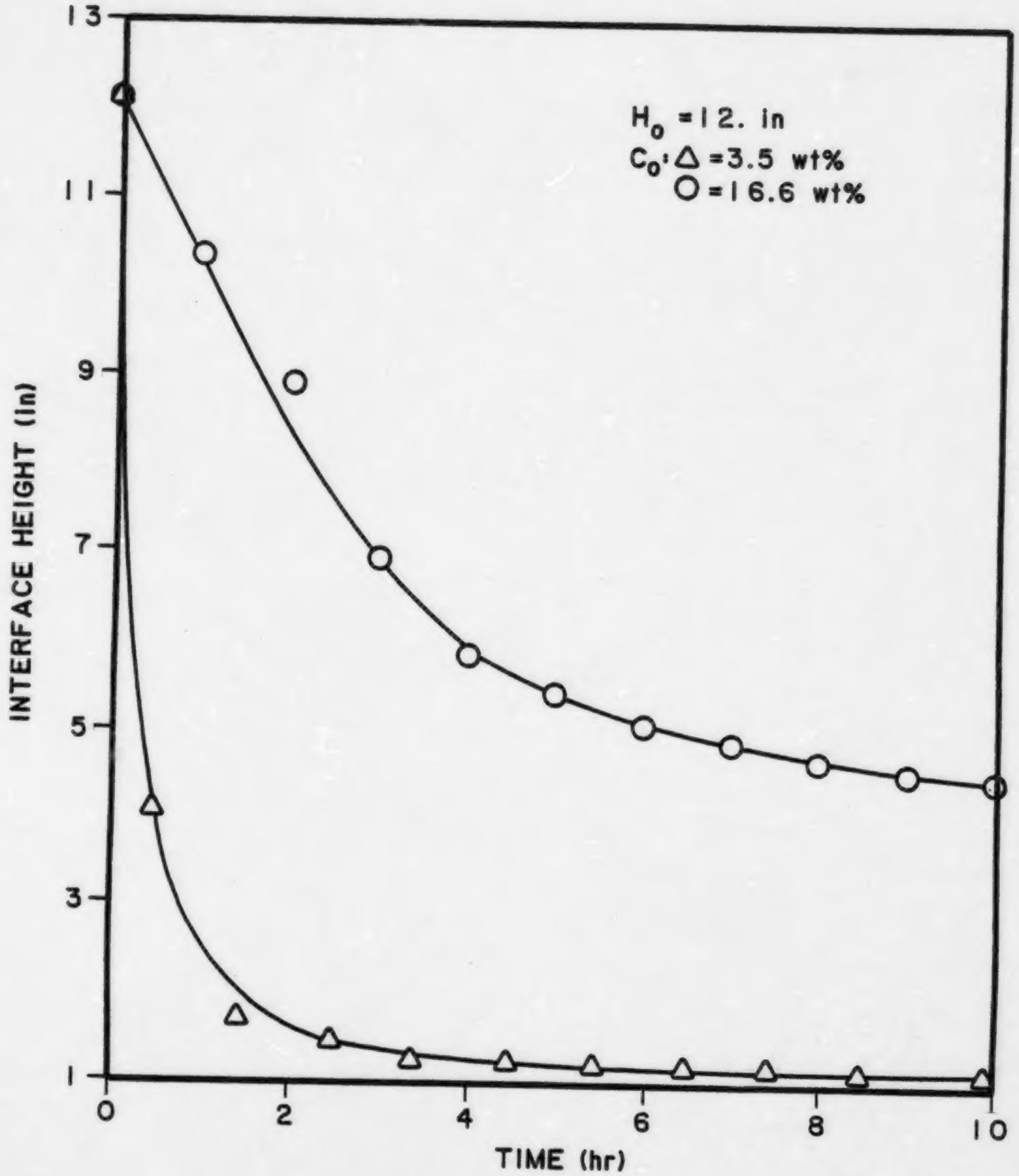


Figure 7.2. Batch settling curves obtained in a short column, H_o = 1 ft.

Table 7.1 Initial Settling Flux for the Batch Settling Tests

Sample	Initial Concentration, C_o (wt%)	Concentration, C $\left(\frac{\text{lb solids}}{\text{ft}^3}\right)$	Initial Settling Velocity, $V_o = -(dH/dt)_o$ (ft/hr)	Flux, $\dot{F}$ $\left(\frac{\text{lb solids}}{\text{ft}^2\text{-hr}}\right)$
1	3.5	2.228	1.3380	2.981
2	5.8	3.742	0.6194	2.318
3	16.4	11.281	0.0926	1.045
4	21.5	15.272	0.0729	1.113
5	25.0	18.167	0.0550	0.999
6	29.85	22.407	0.0390	0.874

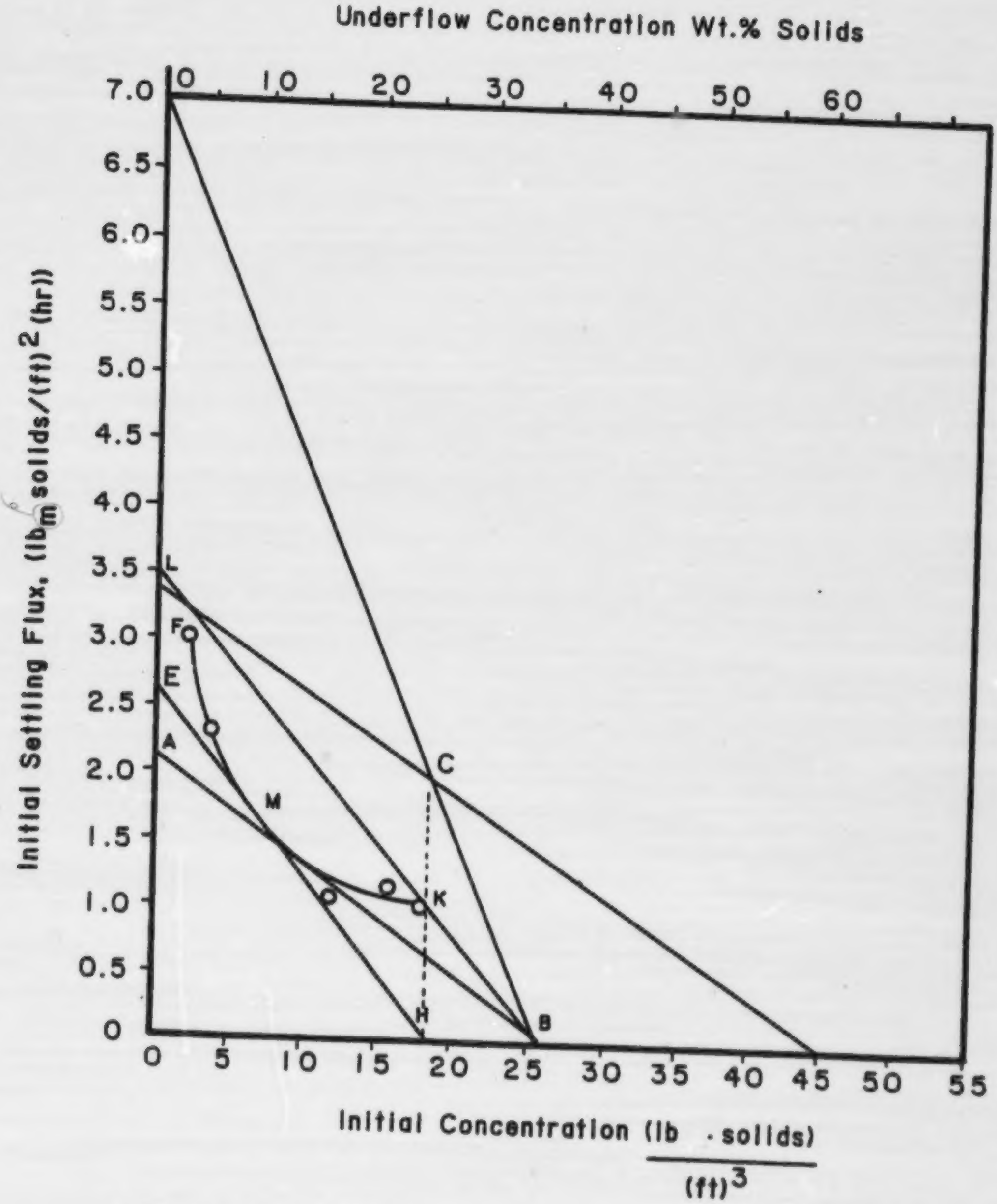


Figure 7.3 Flux Plot for the FGD Wastes of Louisville Gas and Electric

area with the proposed system, the Coe and Clevenger procedure was used in constructing Figure 7.3.

Once a flux plot has been constructed, the critical or limiting flux is determined by drawing a straight line starting at the desired underflow concentration and tangent to the flux curve (In Figure 7.3, an underflow concentration of 35 percent by weight solids has been used for illustration purposes). The flux at the tangent point is considered to be the maximum downward flux of solids possible in a thickener. Given the solids loading (i.e., the amount of solids to be fed per unit of time) for a particular application, thickener area can be computed as given before (Equation 3, Section 5). For a classical type of thickener, if an area less than this computed value is used, then the thickener would eventually fill with solids; the overflow would thus contain solids, and the desired underflow concentration would not be attained.

The degree to which slurries such as FGD wastes can be thickened does not depend on settling area alone. The residence time of solids in the thickener θ_s, also affects the amount of liquid that is separated, or the underflow concentration, C_u. This was observed to be true by Coe and Clevenger (1916) and others (Robert, 1949). With the depth of the compression zone constant, the underflow concentration increases as the residence time of solids increases. The extent to which θ_s affects C_u depends on the physical and chemical characteristics of the feed slurry. This relationship is often explained in terms of the compressive effect exerted by settling solids on underlying layers of solids. Therefore, at a given feed rate, both increased depth of compression zone and increased residence time are necessary for the attainment of higher concentrations in the underflow. As solid-liquid slurries such as FGD wastes settle, a solids concentration is eventually reached at which the solid particles are in contact with one another, and a structural resistance as well as a hydraulic resistance to settling exists. As a result, the amount of solids over-burden increases. Mathematically, this implies that the settling velocity is not a function of concentration (C) alone, but is also a function of the concentration gradient ($\partial C/\partial H$) and possibly of other variables as well (Scott, 1968-a,b, 1970; Fitch, 1966). At present, no simple method exists for predicting the effect of residence time (θ_s) on the dewatering of FGD wastes. However, the solids residence time can be further increased with the use of a clarifier. In the meantime, with the use of a clarifier to prethicken a slurry, the other advantage is that the area requirements for a thickener can be reduced by substituting clarifier area for thickener area, clarifier area being assumed as being roughly one-third as expensive. From Figure 7.3 (tangent line AB), it is estimated that about 0.44 ft^2hr/lb solids are required for a conventional thickener to obtain an underflow concentration of 35 wt%. If the slurry is prethickened by a clarifier to about 25 wt% solids (from H to K), then, only about 0.285 ft^2hr/lb solids are required in the thickener of the combined clarifier-thickener system (line LKB). Actually, the solids flux in the bottom of the thickener of the combined system should be a higher value than that normally possible (due to pre-thickening). Therefore, the breakdown of the floc structure between the clarifier and the thickener could possibly be higher, say, point C, where the solids flux is twice that of point K. In such a case, the area requirement for the thickener in the combined system is about 0.14 ft^2hr/lb solids, which is much lower than that of a conventional thickener.

If the concept of compression settling is valid, then the deeper a layer in a bed of settling solids, the larger the amount of solids overburden, and the more compact, or concentrated, the solids are in the layer. Using this cause-and-effect reasoning, the concept of compression settling was indirectly substantiated in part by batch settling studies. For FGD wastes with different initial solids concentrations and two-settling depths, Figures 7.4, 7.5, 7.6, and 7.7 show: (i) slurry density versus time at certain sludge blanket depths, and (ii) ultimate slurry density profile with depth in a completely settled bed of FGD wastes at certain times. For the minimizing of dynamic effects, these concentration measurements were made after no further change in the solid-liquid interface could be detected. Figures 7.6 and 7.7 seem to indicate that compressive settling did occur. Also, the higher the initial solids concentration (C_0), the less the variation in ultimate slurry density profile.

For FGD wastes having an initial solids concentration higher than 16.4%, no significant variation in solids concentration with bed depth could be detected. The reason for this lack of variation is that the higher the slurry concentration, the stronger the network of solid particles formed (Shirato, 1970), and then, the more resistance to flow of water through the interstices between the solid particles (Shannon and Tory, 1965). Therefore, at higher solids concentrations more overburden is required for the compression of underlying layers of solids. Because of inherent limitations in the measuring device used,

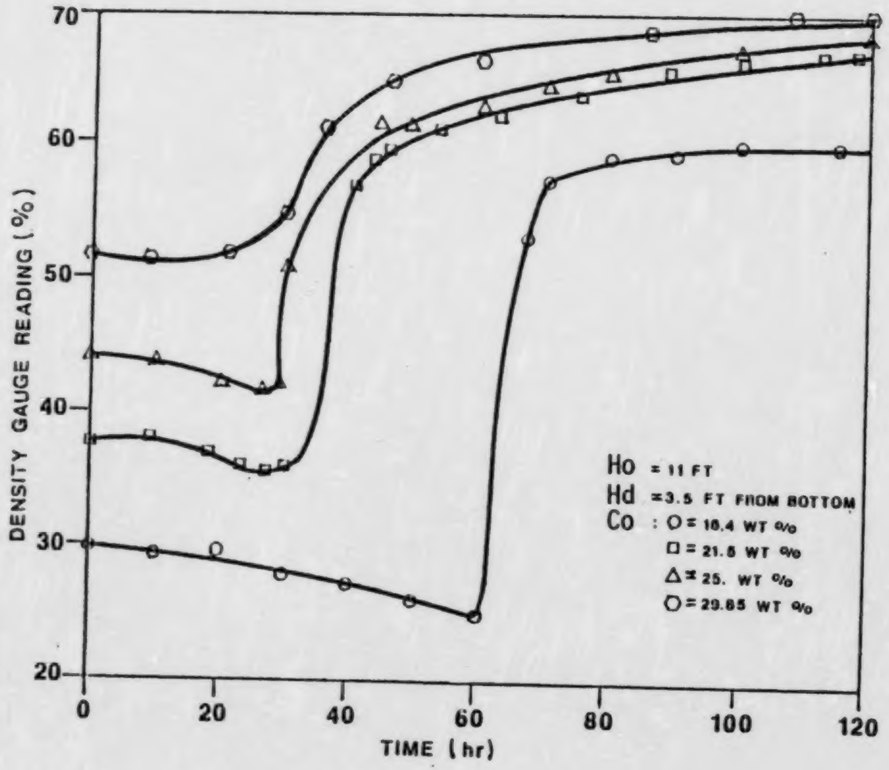


Figure 7.4. The concentration variations of the FGD wastes at a certain point (H_d=3½') during long column tests.

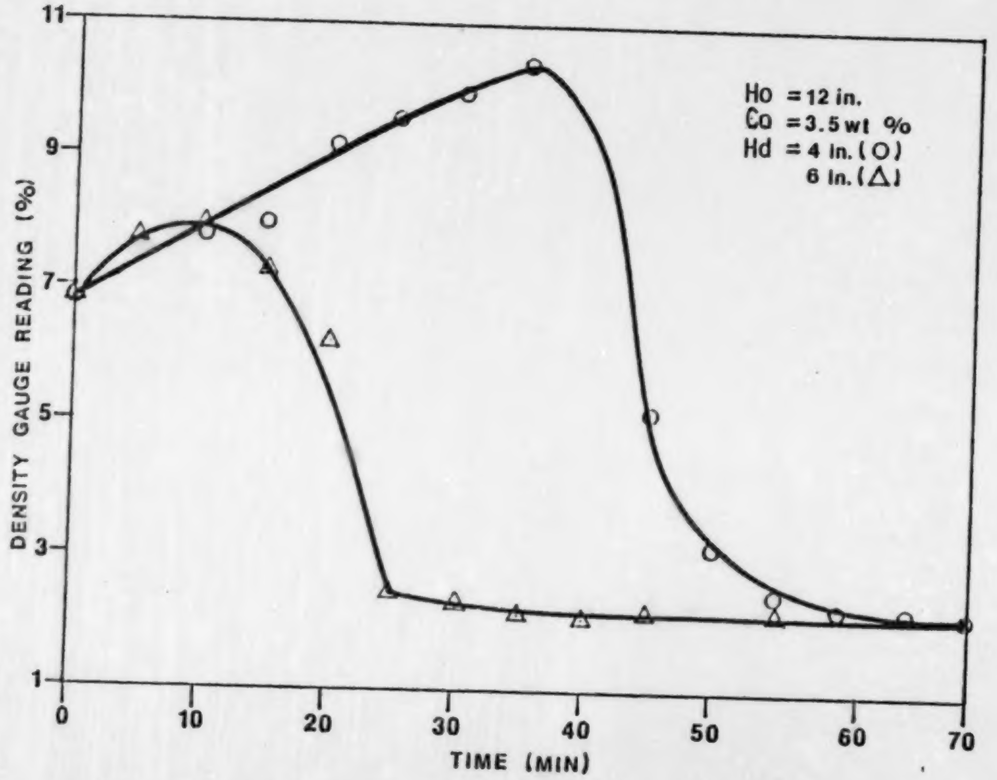


Figure 7.5. The concentration variations of the FGD wastes at a certain point (H_d=4",6") during short column settling tests.

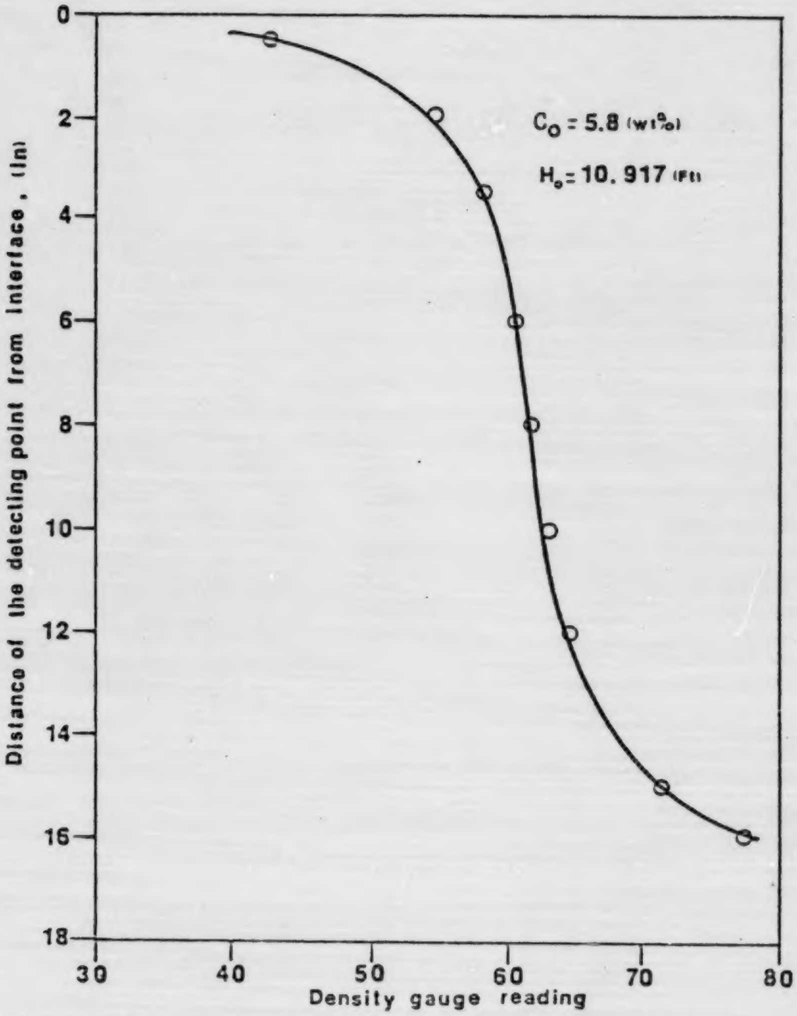


Figure 7.6. Concentration distribution along height after the FGD-waste has been allowed to settle for a long time, C_o=5.8 wt %

Note: Density gauge reading is the direct measure of solids concentration in figures 4 and 5.

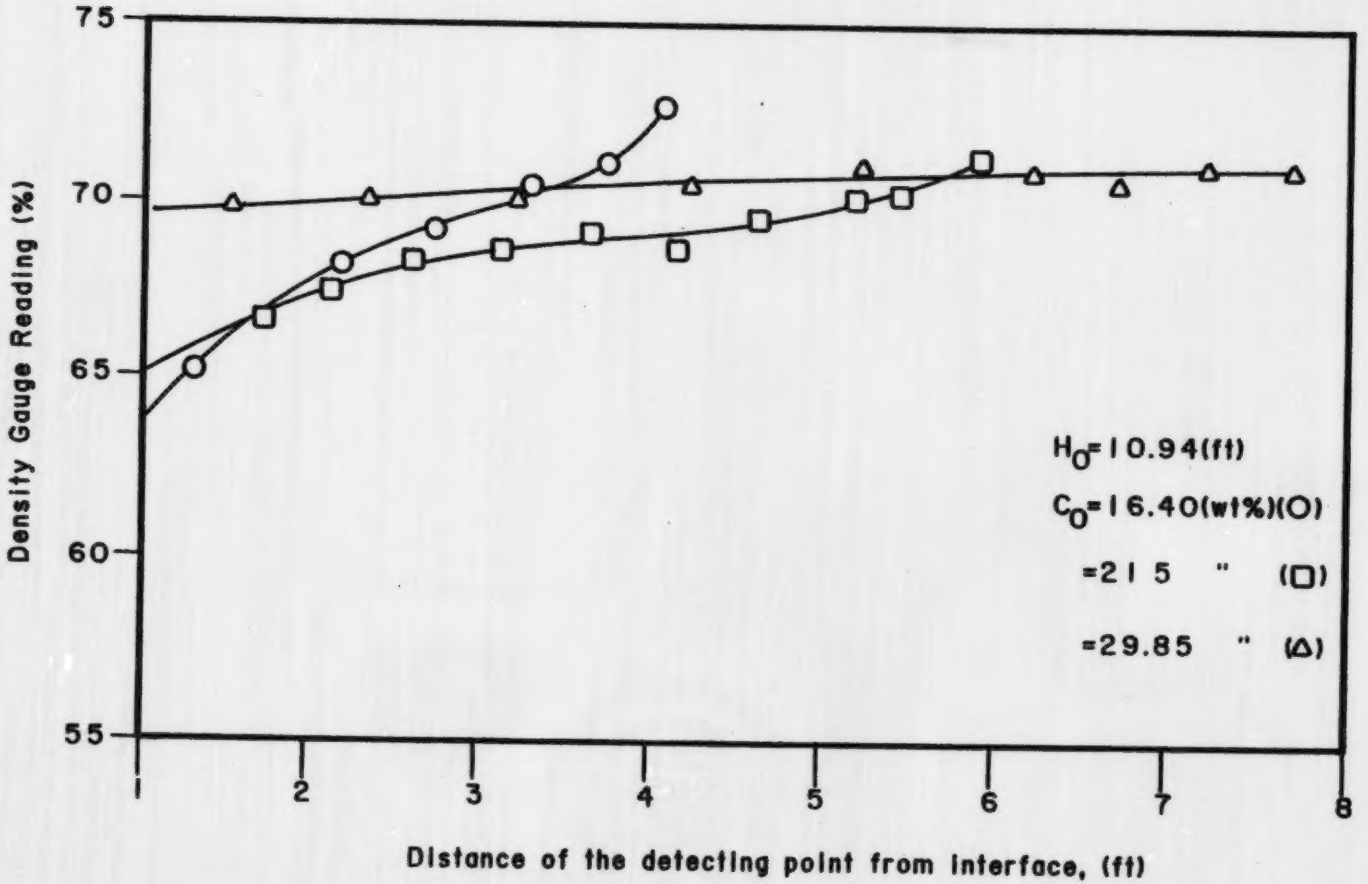


Figure 7.7. Concentration distribution along height after the FGD-waste has been allowed to settle for a long time

the solids concentration profiles, near the top and the bottom of the bed of solids formed, could not be determined accurately, and the expected concentration gradients could have existed in these regions. In other words, the data presented in Figure 7.7 are not conclusive.

In the continuous thickening studies performed, however, the degree to which FGD wastes could be concentrated was observed to depend strongly on the amount of solids accumulation in the thickener, or, the residence time (θ_S). As the amount of solids accumulation in the thickener was increased, the underflow concentration was observed to increase up to some maximum value. After that time, any further accumulation results in an increase in the depth of the compression zone, and a constant settling zone results. Eventually, the excess solids will be eliminated in the overflow. If this happens, it means that the thickener is too shallow or that an impractically high underflow concentration is obtained. This, in turn, results in an imbalance in the mass rate, causing the system to run out of control for a continuous system. Therefore, it is desirable to keep separate the thickener (doing the compression) and the clarifier (doing the prethickening).

Section 8

DESIGN AND OPERATIONAL VARIABLE STUDIES

Exploratory studies were made to determine conceptual criteria for design and operation of a thickener and a clarifier dewatering train to dewater FGD wastes. The thickener and the clarifier were first studied individually. However, the intent of these studies was to determine if and how these units could be used best in combination for dewatering FGD wastes.

8.1 Thickener Studies: The basic variables considered in the thickener studies were solids throughput rate, amount of solids accumulation in the thickener, underflow rate, feed concentration, and location of the feed port. A schematic of the pilot thickener used to make these studies is shown in Figure 8.1. In an attempt to minimize wall effects, a 2' i.d. column was used as the thickener. The feed stock was continuously agitated and circulated, by means of a centrifugal pump, in a 100-gal premixing tank then entered into a 55-gal, constant-head, feed tank. A level controller was fitted to maintain a constant level in the feed tank. The required quantity of feed was drawn from the base of the feed tank, and entered the thickener at a point 9 feet from the base. Clear overflow liquor was siphoned out from the top of the thickener into a premixing tank, where it was mixed with the underflow, thus closing the circuit. The flowrates and concentrations of feed and underflow were measured periodically. FGD

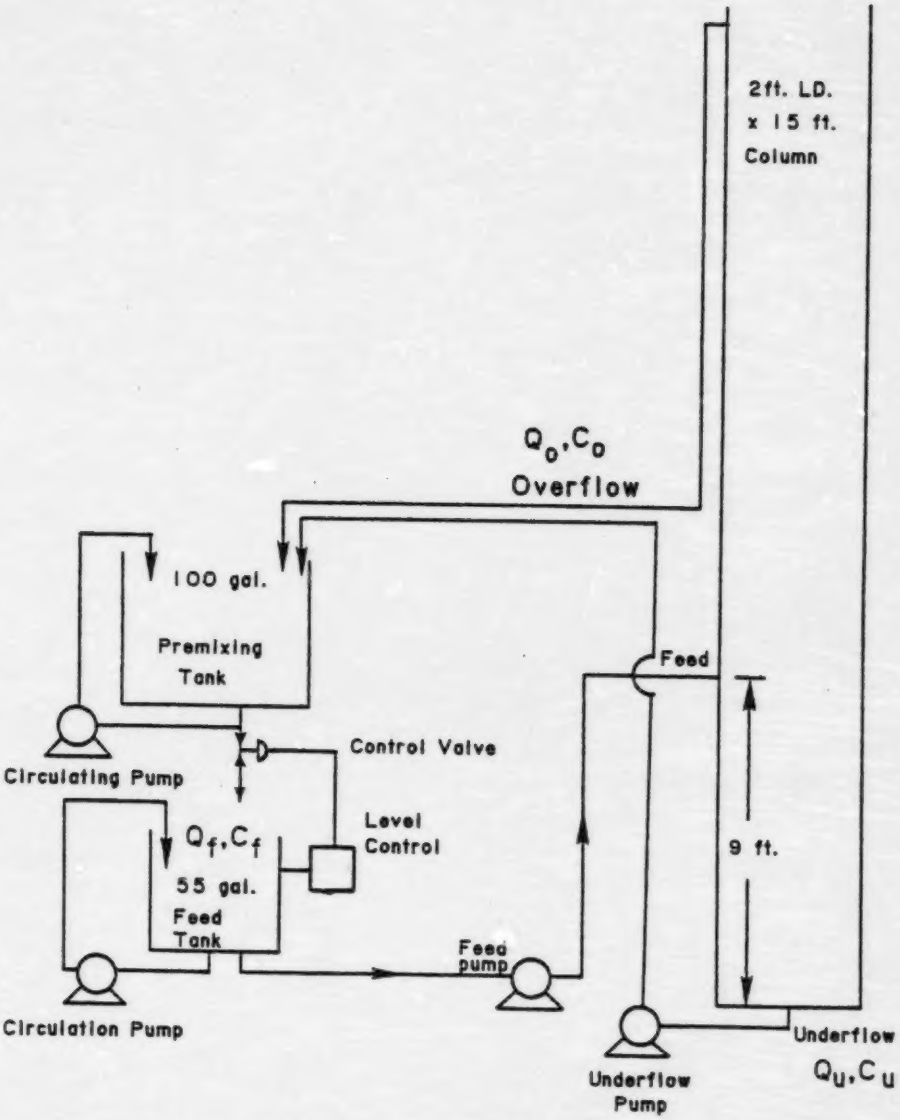

2ft. I.D.
x 15 ft.
Column
Q_o, C_o
Overflow
100 gal.
Premixing
Tank
Feed
Control Valve
Circulating Pump
Q_f, C_f
Level
Control
9 ft.
55 gal.
Feed
Tank
Feed
pump
Circulation Pump
Underflow
Pump
Underflow
Q_u, C_u

Figure 8.1. Schematic of the Pilot Thickener

wastes at high solids concentrations, however, exhibited Bingham fluid behavior, and without stirring, wall effects were unavoidable. Radial concentration profiles at different column heights were measured during testing, and pronounced radial concentration gradients, which acted as flow barriers, were detected. This phenomenon showed that significant wall effects did exist during these studies, causing a reduction in effective settling area and adding an inherent conservatism to the results. Several weeks of continuous operation were required for the attainment of steady-state operation. The experiment was complicated by system upsets caused by the plugging of the lines, valves, pumps, etc. For these reasons, the studies made were limited in that only trends and system sensitivities were established.

The system parameters observed to have the most significant effect on underflow concentration were underflow rate and solids accumulation. As would be expected, when underflow rate was decreased, underflow concentration increased up to some maximum value (Table 8.1). Similarly, as the amount of solids accumulation was increased, thus increasing the solids residence time (θ_s), underflow concentration increased. In the initial studies a low underflow rate and a high amount of solids accumulation were used so as to obtain as high an underflow concentration as possible. After an initial transient start-up period of several weeks, underflow concentration was observed to remain within a range of 35 to 45 percent solids by weight, over a four week operational period.

The feed concentration was maintained at an average of 10.5 wt%. Since it had taken a long operational time to reach approximately steady

Table 8.1 Selected Results of the Pilot Thickener System[a]

Sample	Q_f (ml/min)	C_f (wt%)	Q_u (ml/min)	C_u (wt%)	Q_o (ml/min)	C_o (wt%)
1	4210	10.0	210	42.00	4000	8.00
2	4200	10.0	200	41.00	4000	9.06
3	4200	8.38	200	40.69	4000	7.91
4	950.3	9.46	125	42.43	825.3	6.80
5	953.73	9.00	125	42.84	828.73	7.80
6	951.9	9.46	130	42.27	821.9	7.76
7	904.8	8.85	105	42.74	799.8	8.30
8	800	8.33	105	43.56	800.	6.95
9	879.2	9.12	110	43.74	769.2	6.95

[a] See Figure 8.1

state conditions, no attempts were made to adjust feed conditions, so as to improve overflow clarity. In this experiment, the overflow concentration averaged about 8 wt.%. Unless the overflow concentration was at its maximum value and the solids feed matched the underflow solids draw-off rate, that is, the solids accumulation in the column stopped (or the compression zone stopped building up), overflow clarity was assumed to be attainable by using a clarifier in conjunction with a thickener. This idea was proved to be true in subsequent studies (see Section 10). The only measurements made for this system were (i) the maximum underflow concentration attainable, and (ii) the change in underflow concentrations versus the feed rates.

The settling tests were performed daily as soon as it was possible to operate the system in a desirable fashion. It was found that constant underflow concentrations could not be maintained because the underflow line frequently plugged because of its small size (3/8" I.D.). Therefore, constant feed concentrations were also difficult to maintain. For these reasons, true steady state operation could not be maintained over an extended period (i.e., for a period of time longer than the residence time of the solids within the thickener). However, these tests did serve to illustrate that extremely high underflow concentrations could result with simple gravity sedimentation.

Under "superficial" steady-state operation, the underflow concentration obtained with FGD wastes was 50 ± 2 wt. percent during the last week of continuous operation. The physical and chemical properties of FGD wastes, therefore, are such that solids residence time (θ_s), and compressive settling are important in the dewatering. It was very difficult to quantify the effect of θ_s, or solids accumulation, on underflow concentration under a flooded condition (i.e., the overflow was turbid). Consequently, the amount of solids in the thickener could not be gauged simply by measuring the height of the blanket of thickened solids in the bottom. In addition, it was almost impossible to interpret concentration profile measurements in this thickener, because of radial concentration gradients.

To obtain the large amount of solids accumulation for these initial studies, solids were fed gradually into the thickener. As more solids accumulated in the thickener, underflow concentration increased gradually to a relatively constant level of about 35% to 45% solids by weight. The underflow rate, at this time, was adjusted periodically so that one might try to maintain steady state operation. This mode of operation was continued for a four-week period, at which time this experiment was terminated, and the thickener was drained and cleaned for another experiment.

During continuous operation of the thickener over an eight-week period, underflow concentration (Cu) remained at 43 ± 2% as shown in Table 8.1. Underflow concentration was shown actually to be affected by underflow draw-off rate, although the differences manifested in this experiment were not very great. These differences could be explained by large fluctuations in feed rates and feed concentrations. Also, the underflow concentration was recorded to be as high as 50 wt% only once or twice during this two-month period. The large difference in underflow concentrations obtained under the same conditions, could be attributed only to the difference in solids accumulation, and, presumably, solids residence time.

8.2 Clarifier Studies: For the evaluation of the limitations of the use of a laminar-type clarifier to dewater FGD wastes, a set of experiments were made with the use of the apparatus shown in Figure 8.2. In this experiment, a 4' long, 1'-7-1/2" ID glass tube was used as a clarifier instead of the laminar-type, where inclined plates were put in the tube to provide additional internal surface area. The feed rates and the concentrations were kept constant. During the experiments the following variables were considered: settling area, angle of inclination, solids loading, and location of the feed port. As seen from Table 8.2, the additionally provided surface area contributed significantly to the separation efficiency or underflow concentration. On the other hand, the longer the tubular clarifier, the greater the residence time and area. Thus, a greater separation efficiency was obtained (Table 8.3). Similarly, the location of the level of the solid-liquid interface in the clarifier was observed to be important: the higher the interface the greater the residence time and the separation efficiency (Table 8.4). Also, the best location of the feed port (near the top rather than the bottom of the clarifier) was observed to have a significant beneficial effect on the performance, since the entering fluid would not disturb the stationary bulk of settling fluid (Table 8.5). Often, in conventional systems, the feed is introduced at the bottom of the clarifier. It is interesting that, in the most recently released product bulletin of Parkson's Lamella thickener, the feed is introduced no longer at the bottom of the settling plates but at

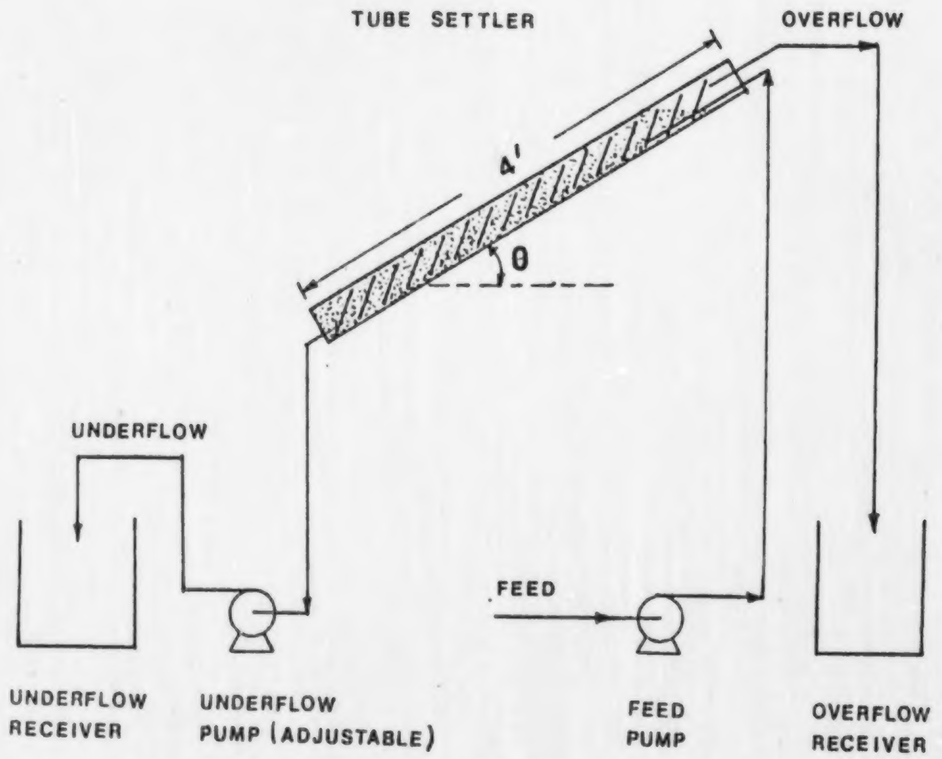

TUBE SETTLER
OVERFLOW
4'
θ
UNDERFLOW
FEED
UNDERFLOW RECEIVER
UNDERFLOW PUMP (ADJUSTABLE)
FEED PUMP
OVERFLOW RECEIVER

Figure 8.2. Tube settler with inclined plates.

Table 8.2 Effect of Inclined Plates on Separation Efficiency and Underflow Concentration for a Tube-type Clarifier

Case		Underflow Concentration (wt%)	Separation Efficiency, $(Q_{oc}/Q_{uc})(100\%)$
1.	A	15.1	48.78
	B	13.2	41.70
2.	A	19.7	56.2
	B	16.2	50.3

Feed concentration = 10 wt%
Feed flow rate = 3. l/hr.
Angle of inclination = 30° (tube)
= 22° (plates to the tube)
A =with inclined plates
B = without plates
Location of interface = top (case 1)
= middle (case 2)
Clear overflow

Table 8.3 Effect of Tube Length on Separation Efficiency and Underflow Concentration for a Tube-type Clarifier

Tube Length (ft)	Underflow Concentration (wt%)	Separation Efficiency, $(Q_{oc}/Q_{uc})(100\%)$
4	16 14 avg	38 34 avg
4	12	31
8	18	53
8	20 18 avg	52 52 avg
8	16	51

Feed concentration = 10 wt%
Feed flow rate = 3. l/hr.
Angle of inclination = 30° (tube)
Solid-liquid interface @ 5" from top
Clear overflow

Table 8.4 Effect of Location of Solid-Liquid Interface on Separation Efficiency and Underflow Concentration for Tube-type Clarifier

Location of Solid-Liquid Interface	Underflow Concentration (wt%)		Separation Efficiency $(Q_{oc}/Q_{uc})(100\%)$	
5" from top	18	18 avg	56	56 avg
5" from top	18		56	
37" from top	16	17 avg	40	39 avg
37" from top	17		37	

Feed concentration = 10 wt%
Feed flow rate = 3. 1/hr.
Angle of inclination = 30°
Tube length = 8 ft.
Feed port @ 2" below solid-liquid interface
Clear overflow

Table 8.5 Effect of Location of Solid-Liquid Interface on Separation Efficiency and Underflow Concentration for Tube-type Clarifier

Location of Feed Port	Underflow Concentration (wt%)	Separation Efficiency $(Q_{oc}/Q_{uc})(100\%)$
1 ft. below the top	16	57
Bottom	12	34

Angle of inclination = 30°
Feed concentration = 8.8 wt%
Feed flow rate = 6.5 l/hr.
Solid-liquid interface @ 20" from top
Modified tube with additional internal surface area
Tube length = 4 ft.
Clear overflow

some higher point in the unit (Parkson Corporation, 1977). More importantly, it is better to feed at the bottom than at the top of the cross-section and concurrently with the downward flowing solids. Such a location will reduce the extent of remixing of the separated, upward flowing clear water with the feed or the separated, downward flowing solids. The angle of inclination (θ in Figure 8.2) observed to give the highest separation efficiency was about 30° (Table 8.6). The higher the angle of inclination, the less the projected separation area. However, a lower angle should give a higher separation efficiency. But, as a result of lowering the angle, the draw-off rate of the concentrated solids was decreased and more solids accumulation in the clarifier resulted. Gradually, the tubular clarifier was occupied by the solids, and the separation area decreased. Therefore, the tube angle versus separation efficiency plot should give a maximum at a certain angle for a specific slurry. This optimum point would be different for different kinds and different concentrations of slurries.

The above results showed that the separation efficiency increases almost linearly with increases in settling area, but underflow concentration increases only slightly. It is also true that when the solids loading is decreased, the separation efficiency increases at a certain rate (as shown in Figure 8.3). These results imply that, although it can separate liquids from solids faster than a conventional clarifier/-thickener, a laminar-type clarifier does only a partial job of thickening. Actually, the FGD wastes tested could be thickened only to concentrations of 15 to 25% with the use of a laminar-type clarifier alone for 10 wt% initial feed concentration and such separating conditions.

Table 8.6 Effect of Angle of Inclination on Separation Efficiency and Underflow Concentration for Tube-type Clarifier

Angle of Inclination (degrees)	Underflow Concentration (wt%)	Separation Efficiency $(Q_{oc}/Q_{uc})(100\%)$
45	13	34
30	14	41
25	14	41

Feed concentration = 10 wt%
Feed flow rate = 6 l/hr.
Solid-liquid interface @ 5" from top
Modified tube with additional internal surface area
Tube length = 4 ft.
Clear overflow

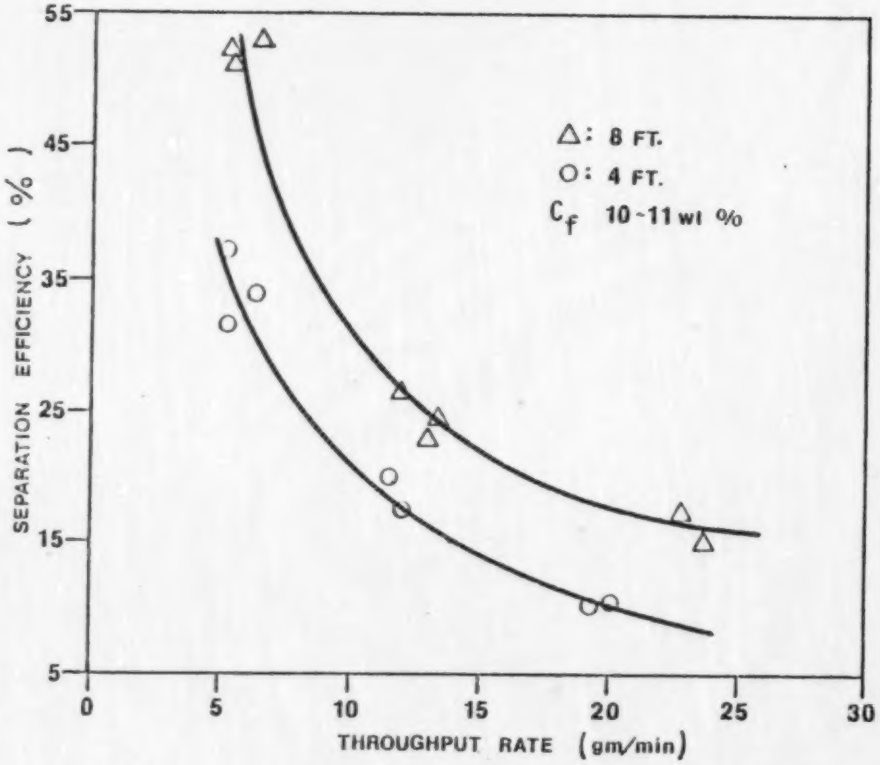


Figure 8.3. Effect of feed rate on separation efficiency of the tube settler.

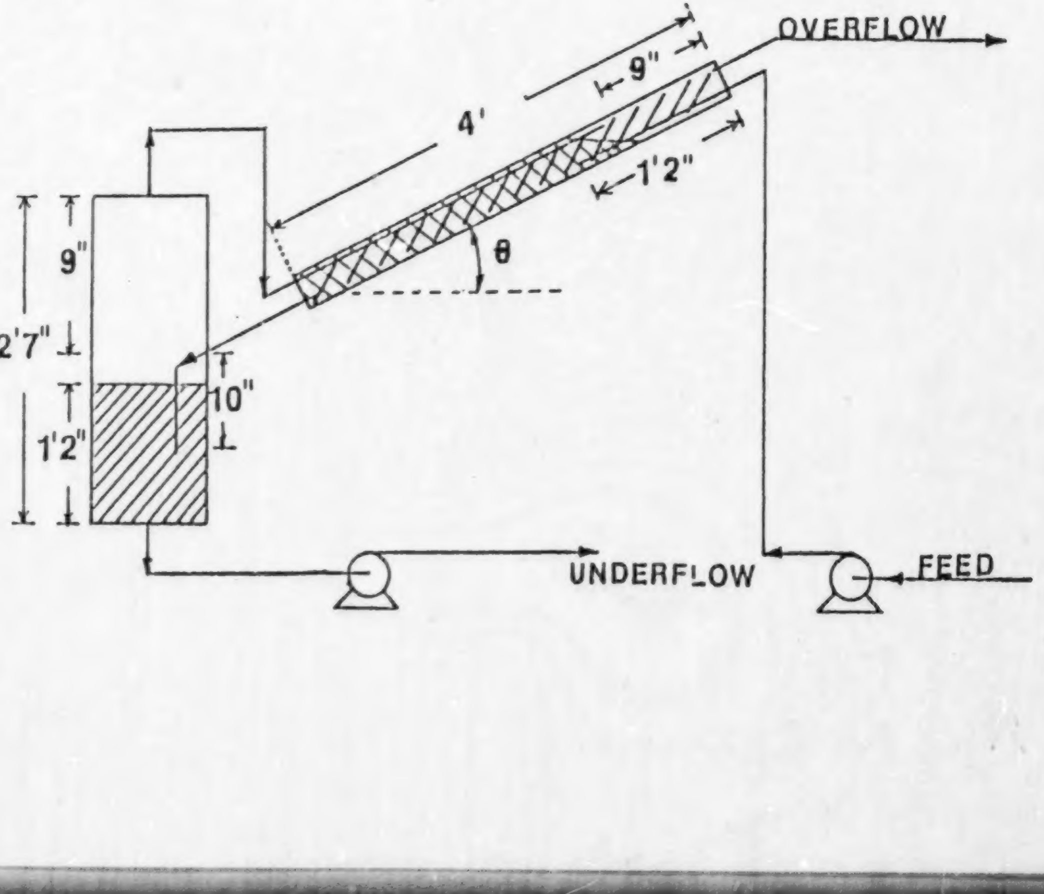
However, this range of underflow concentrations is above the value of the tangent point in the solids flux plot given in Figure 7.3 and indicates that a prethickening of the feed using a laminar-type clarifier should serve not only to increase solids residence time but also the critical downward solids flux in the thickener.

Section 9

DESCRIPTION OF THE PROPOSED DEWATERING SYSTEM

Using the results and speculations discussed in the previous section, the initial version of the new dewatering system was constructed as shown in Figure 9.1. It was similar to the system shown in Figure 8.2 and uses a tube settler, which is connected to a 3-1/8" ID, 2' 7" high cylindrical receiver to serve as a thickener. The water coming out of the receiver flowed into the settler and combined with the upward flowing separated water. It was found that it was impossible to keep the interface at a certain level in the tube settler because interface level in the tube was higher than that in the receiver. So, after the system was run for a certain time, the receiver was fully filled with thick slurry, and the overflow line was also plugged by the thick solids. Therefore, the function of the receiver was kept at a minimum. The concentration of the underflow from the receiver was almost the same as that from the tube settler. Actually, this unit was very similar to the conventional laminar-type clarifier except that the separated water from the receiver could bypass the settling tube and enter the clear overflow. This unit did perform the expected roles for both clarifier and thickener. However, the clarifier and the thickener were not decoupled.

FIGURE 9.1. Tube settler with cylindrical receiver


OVERFLOW
9"
4'
1'2"
θ
9"
2'7"
10"
1'2"
UNDERFLOW
FEED

In order to achieve clarification in a clarifier and compression in a thickener separately, it was decided that the underflow line from the tube settler to the receiver should be interrupted. Thus, the idea of using recycle was conceptualized. A flow diagram of the more desirable gravity sedimentation system proposed for the dewatering of FGD wastes is given in Figure 9.2. A classical thickener and a laminar-type clarifier are used in combination and provided with a recycle stream, the role of the clarifier being to prethicken the FGD wastes as well as to clarify the overflow and that of the thickener, to thicken the wastes as much as possible. The use of a laminar-type clarifier and a thickener in combination is by no means a novel concept in thickener technology. In conventional systems, however, the clarifier is usually mounted above the thickener, and the thickener is usually just a relatively small sump tank. In the proposed system (Fig. 9.2), the thickener area can be as large or as small as desired, and the recycle stream can be fed into the thickener at any desired height. In other words, in the proposed system the clarification and the thickening function of the system are decoupled as much as physically possible. In addition, unlike conventional systems, the recycle rate can be controlled. Consequently, it is possible to have a fine operational control of the dewatering systems that is very valuable.

Control of the recycle rate from the clarifier to the thickener is important in that it affects both thickener and clarifier performance. As an aid in illustrating the importance of controlling recycle rate, steady-state computer analyses of the proposed system and of a conventional system were made. The two systems, for simplicity, were

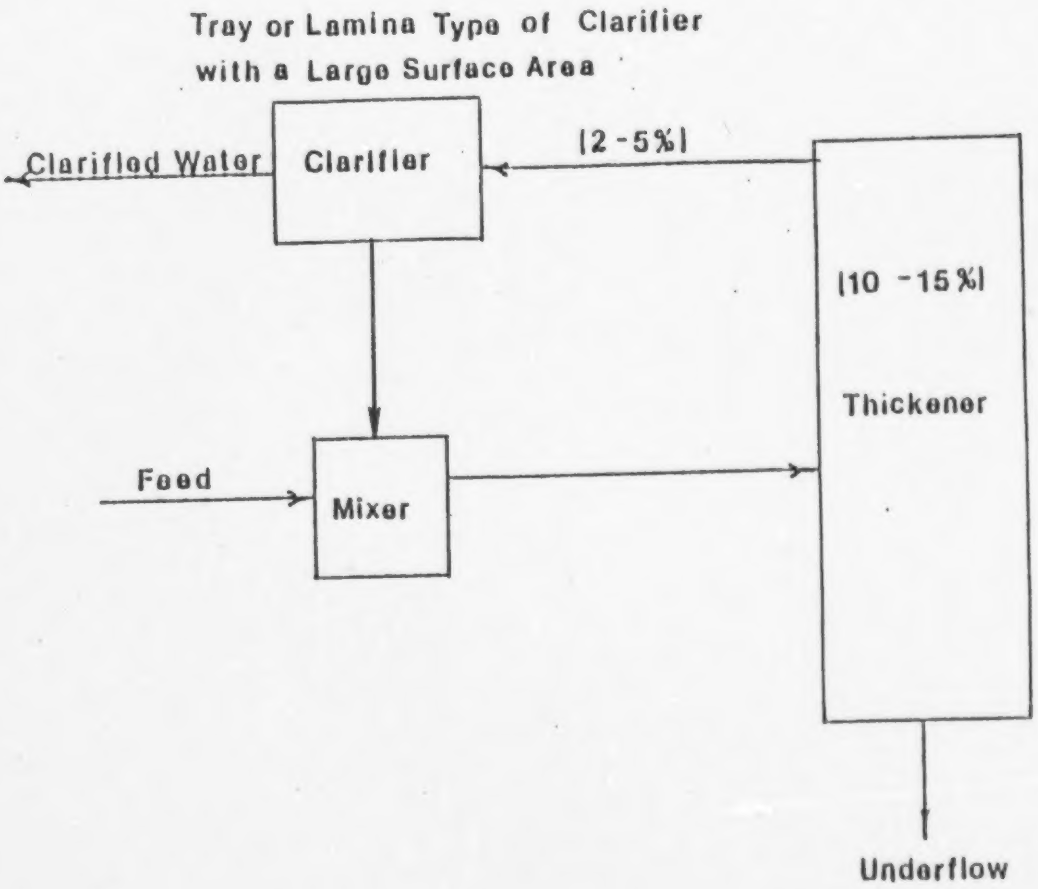

Tray or Lamina Type of Clarifier
with a Large Surface Area
Clarified Water
Clarifier
(2 - 5%)
(10 - 15%)
Thickener
Feed
Mixer
Underflow

Figure 9.2. Flow Diagram of Proposed Initial Treatment System

simulated by the use of the discrete unit operations and the flow scheme shown in Figure 9.3.

The top portion of the thickener was simulated by the assumption that the mixing and the separation of the different streams entering this section are independent actions superimposed together. In other words, the upper portion of the thickener is simulated as a mixer because some mixing of the different streams occurs, and a splitter because some separation of water or thickening also occurs. In these steady-state analyses underflow concentration was held constant, and a steady-state material balance was performed for specific clarifier underflow concentrations, recycle flow rates, and separation efficiency at the top portion of the thickener (i.e., splitter efficiency, defined as C_{us}/C_{os}). Some results of the steady-state analyses are given in Figures 9.4, 9.5, 9.6, and 9.7.

Apparently, as the ratio of the recycle rate to the feed rate (Q_{uc}/Q_f) is increased, the volumetric flow rate of slurry of solids into the clarifier (Q_{os}) where the prethickening occurs is increased. In other words, the recycle probability of slurry in this section is largely increased. Thus, the total residence time of solids in the clarifier is increased. As a result, at constant feed rate, feed concentration, and a unit time, the total separated water per unit mass of solids in this section will be increased. So the underflow concentration from the splitter (C_{us}) could be increased by decreasing Q_{us}.

Although the total function of the clarifier can be increased by increasing the recycle rate (Q_{uc}), the underflow concentration from the

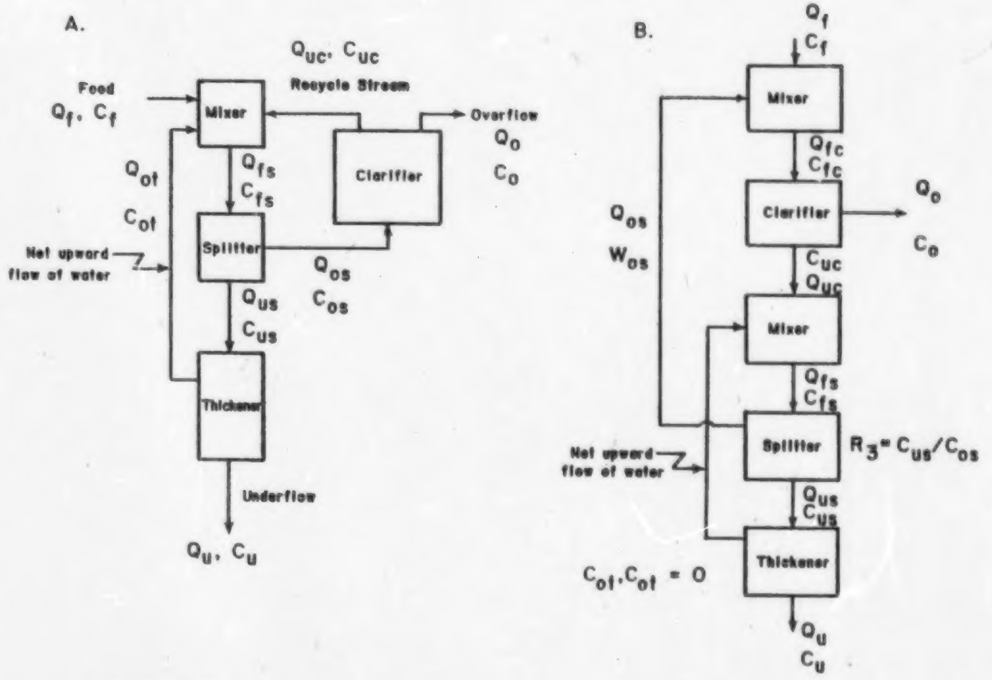

A.
Feed
Q_f, C_f
Mixer
Q_{uc}, C_{uc}
Recycle Stream
Overflow
Q_o
C_o
Clarifier
Q_{fs}
C_{fs}
Q_{ot}
C_{ot}
Net upward flow of water
Splitter
Q_{os}
C_{os}
Q_{us}
C_{us}
Thickener
Underflow
Q_u, C_u
B.
Q_f
C_f
Mixer
Q_{fc}
C_{fc}
Clarifier
Q_o
C_o
Q_{os}
W_{os}
C_{uc}
Q_{uc}
Mixer
Q_{fs}
C_{fs}
Net upward flow of water
Splitter
$R_3 = C_{us}/C_{os}$
Q_{us}
C_{us}
Thickener
$C_{ot}, C_{ot} = 0$
Q_u
C_u

Figure 9.3. Schemes of discrete operations used to simulate the proposed dewatering system

A. Proposed system configuration

B. Conventional system configuration

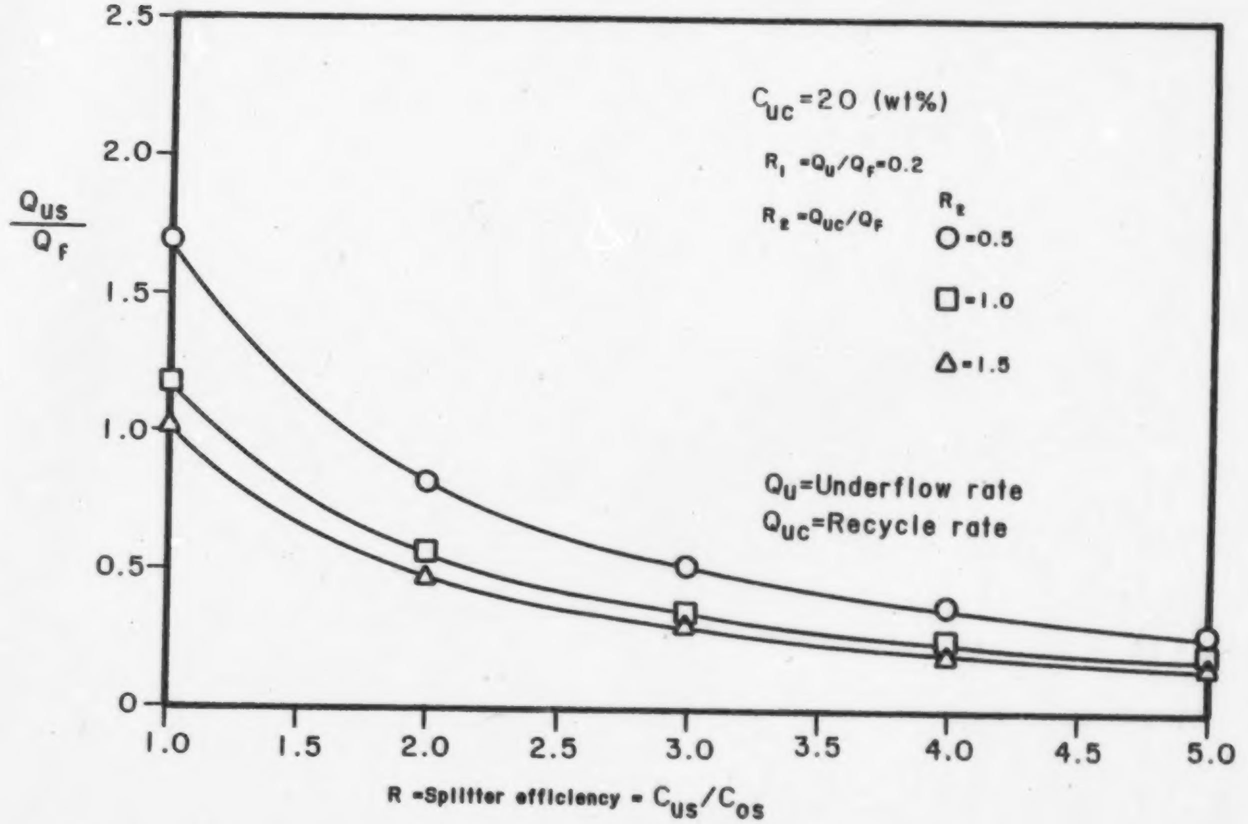


Figure 9.4. Variation of the underflow rate Q_{us} from splitter to the thickener for the proposed system configuration.

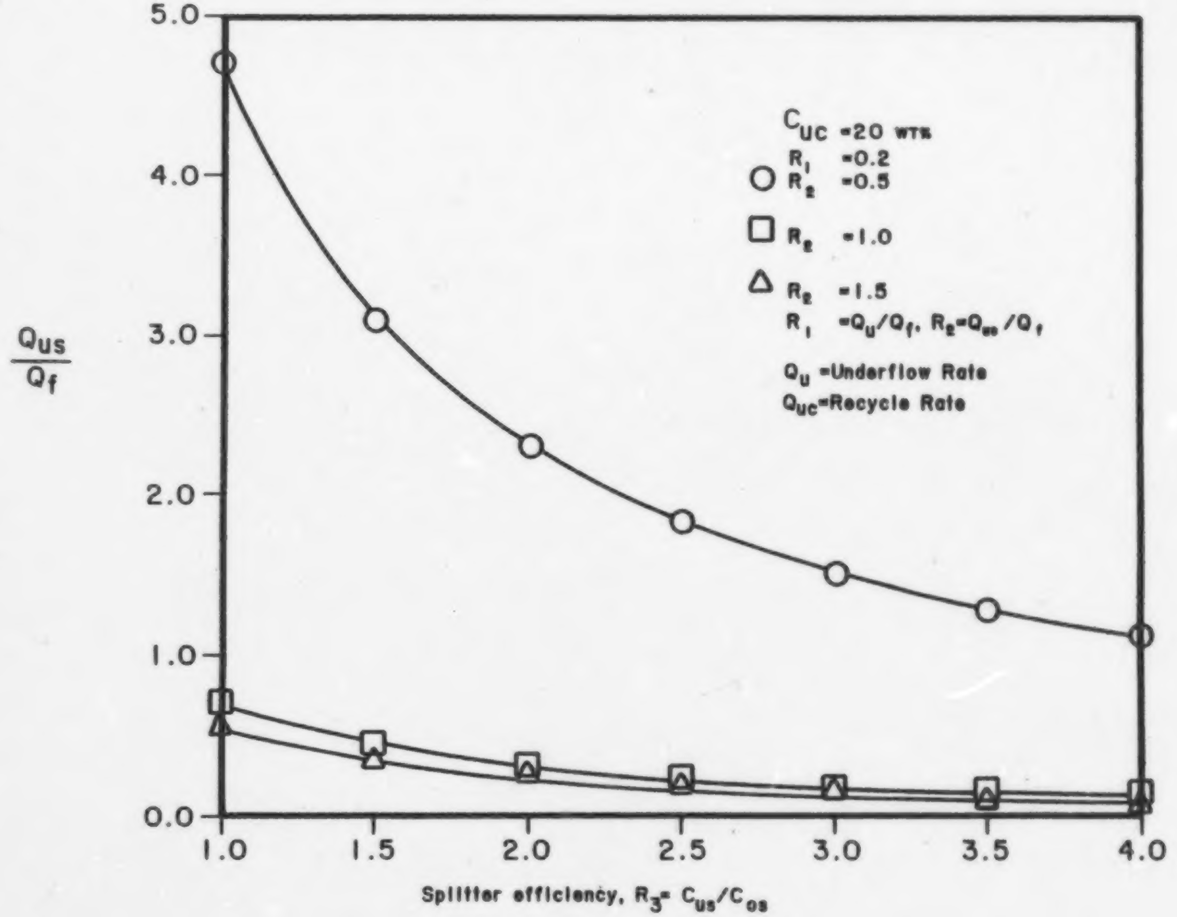


Figure 9.5. Variation of the underflow rate from splitter to the thickener for the conventional system configuration

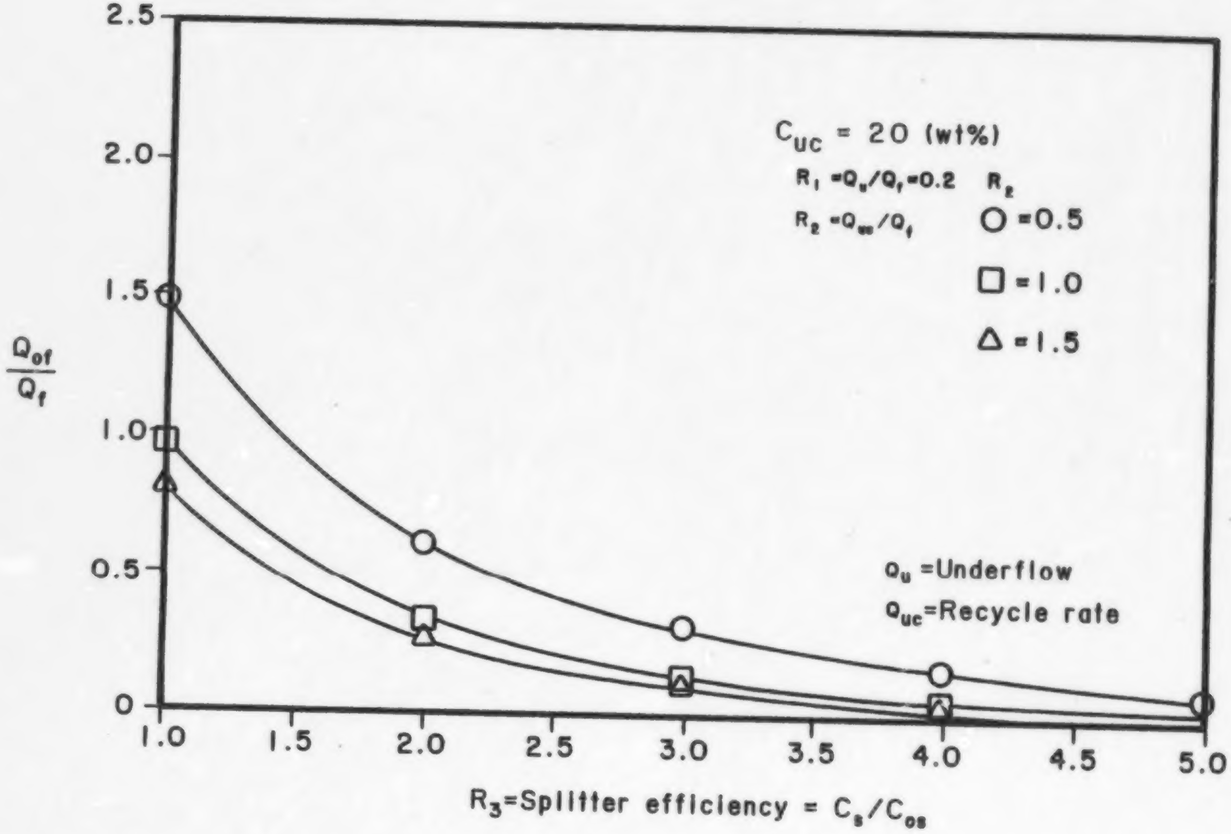


Figure 9.6. Variation of the overflow rate Q_{of} from thickener for the proposed system configuration.

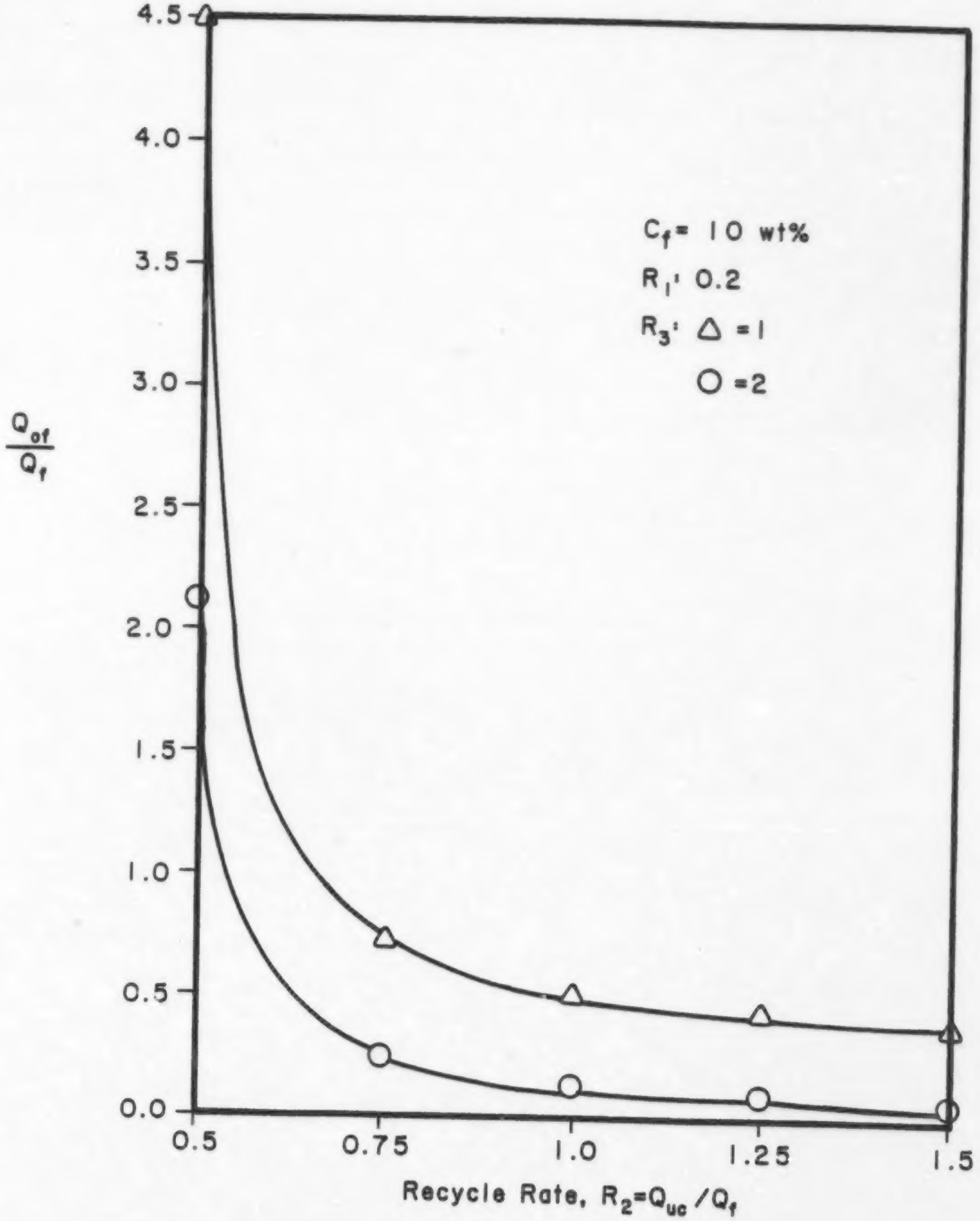


Figure 9.7. Variation of the Overflow Rate Q_{ot} from Thickener for the Conventional System Configuration.

clarifier (C_{uc}) could be decreased. The concentration difference between feed stream and recycle stream will be decreased because the greater portion of slurry is recycling in the closed circuit (mixer → splitter → clarifier). The case is the same for the concentration from the mixer (C_{fs}) and the one to the clarifier (C_{os}). As the recycle rate is further increased, the underflow concentration from the splitter (C_{us}) will go up to some maximum value and will then decrease. In other words, at the constant separation area available in the clarifier, if the recycle rate is very large compared with the feed rate (Q_f), the addition of the fresh feed and the net upward flow of water to the recycle stream is too small to be counted in the recycle stream.

Then:

$$Q_{uc} \doteqdot Q_{fs} \doteqdot Q_{os} \qquad (14)$$

$$C_{uc} \doteqdot C_{fs} \doteqdot C_{os} \qquad (15)$$

and the overflow of clarifier (Q_o) approaches zero. In such a case,

$$C_f \doteqdot C_{fs} \doteqdot C_{os} \doteqdot C_{uc} \doteqdot Cus \qquad (16)$$

$$Q_{us} = Q_{fs} - Q \doteqdot Q_f \qquad (17)$$

Since mixing predominates in the clarifier in such a case, it is impossible to have settling, and it seems as if the functions of the mixer, the splitter and the clarifier are of no use unless there is infinite separation area in the clarifier. This case behaves the same as a conventional thickener with feed (Q_{us}) in, and overflow (Q_{ot})

and underflow (Q_u) out. Consequently, the recycle rate is a very important operation factor. The optimum point, which depends on the feed rate, feed concentration, the separation area available in the clarifier, the quality of the slurry, etc., should be attained by a qualitative test of the system. In addition to causing a decrease in C_{us}, the increase in the concentration of the feed to the clarifier, by augmenting the recycle rate, also causes a decrease in the separation efficiency of the clarifier, and the desired overflow clarity may not be attained unless there is an increase in the separation area in the clarifier. In effect, then, more or less load can be placed on the clarifier by a simple adjustment of the recycle rate.

The volumetric flow of slurry to the lower section of the thickener (Q_{us}) depends on the amount of separation that occurs at the top of the thickener (i.e., splitter efficiency, C_{us}/C_{os}, as Shown in Figure 9.4) as well as on recycle rate. In other words, the amount of mixing of water separated in the bottom of the thickener with the feed and the recycled wastes affects the split between the flow of these streams to the clarifier and that to the bottom of the thickener. The amount of mixing or separation that results in this section would depend on many mechanical design factors and operational conditions. An advantage of the proposed system is that the recycled waste can be fed into the thickener below the feed port and zone of lower concentration, whereas for conventional systems the bottoms of the clarifier are fed into the very top of the thickener. Also, compared with the conventional system (Figure 9.5), at higher recycle rates, the proposed system has a better splitter efficiency than the conventional system. On the other hand, to obtain high underflow concentrations, the amount of water that has to be displaced in the bottom of the thickener is substantial (Q_{ot} in Figure 9.6), and the remixing of this separated water with the clarifier underflow (or recycle wastes) and the influent wastes can indeed have a significant effect on overall separation efficiency.

Feed concentration and solids loading were also observed to be important operation variables. In order to obtain steady state operation and a clear overflow, underflow rate has to be increased or decreased with changes in the amount of solids fed into the thickener. As shown earlier, underflow concentration is very sensitive to changes in underflow rate or solids residence time (θ_s). As an illustration of the interdependence of solids loading, underflow rate, and underflow concentration, a set of experiments was conducted using the apparatus shown in Figure 9.8. While the feed rate and concentration (i.e., the solids loading) were kept constant, underflow rates (Q_u) were varied, as shown in Figures 9.9 and 9.10. At high underflow rates, underflow concentration dropped to essentially that of feed. In other words, the feed was flowing directly out of the underflow without any occurrence of settling. At low underflow rates, on the other hand, overflow concentration approached that of the feed, and underflow concentration increased to a maximum value of about 25%. This increase in underflow concentration with decrease in underflow rate is assumed to be due to the resulting increase in the residence time of the solids in the unit. For the attainment of a clear overflow, Q_u has to be greater than about 110 ml/min, as shown in Figure 9.9. When Q_u was less than 110 ml/min, and the feed concentration was about 10%, C_u was very sensitive to the

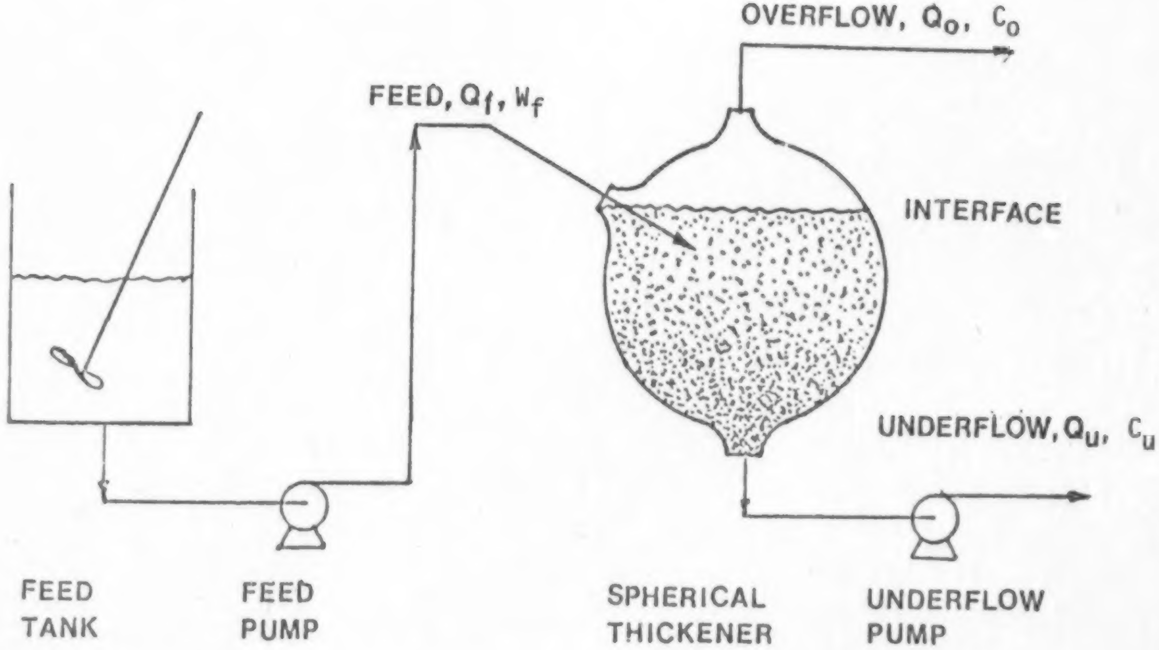

OVERFLOW, Q_o, C_o
FEED, Q_f, W_f
INTERFACE
UNDERFLOW, Q_u, C_u
FEED TANK
FEED PUMP
SPHERICAL THICKENER
UNDERFLOW PUMP

Figure 9.8. SPHERICAL SETTLER

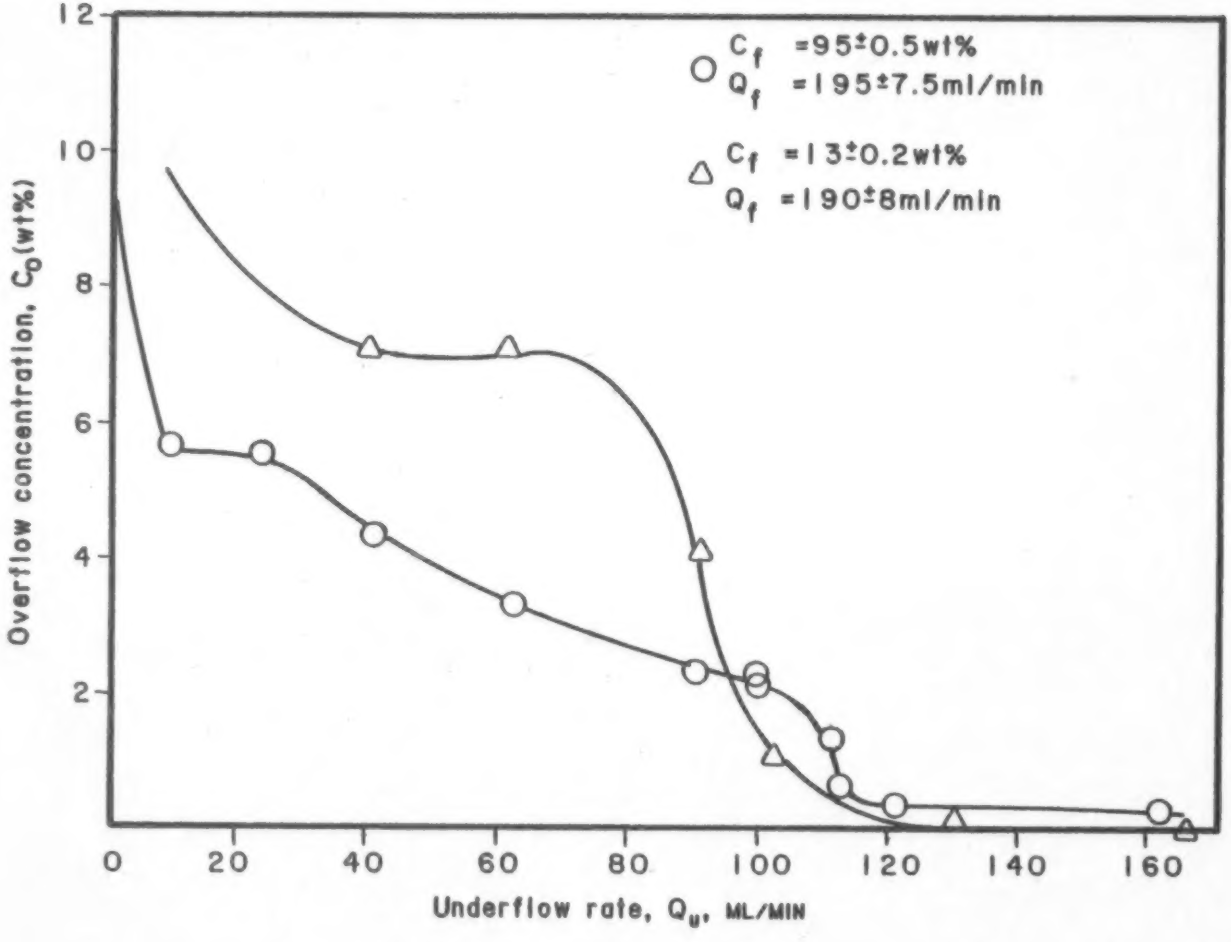


Figure 9.9. Spherical settler: effect of underflow rate on overflow concentration.

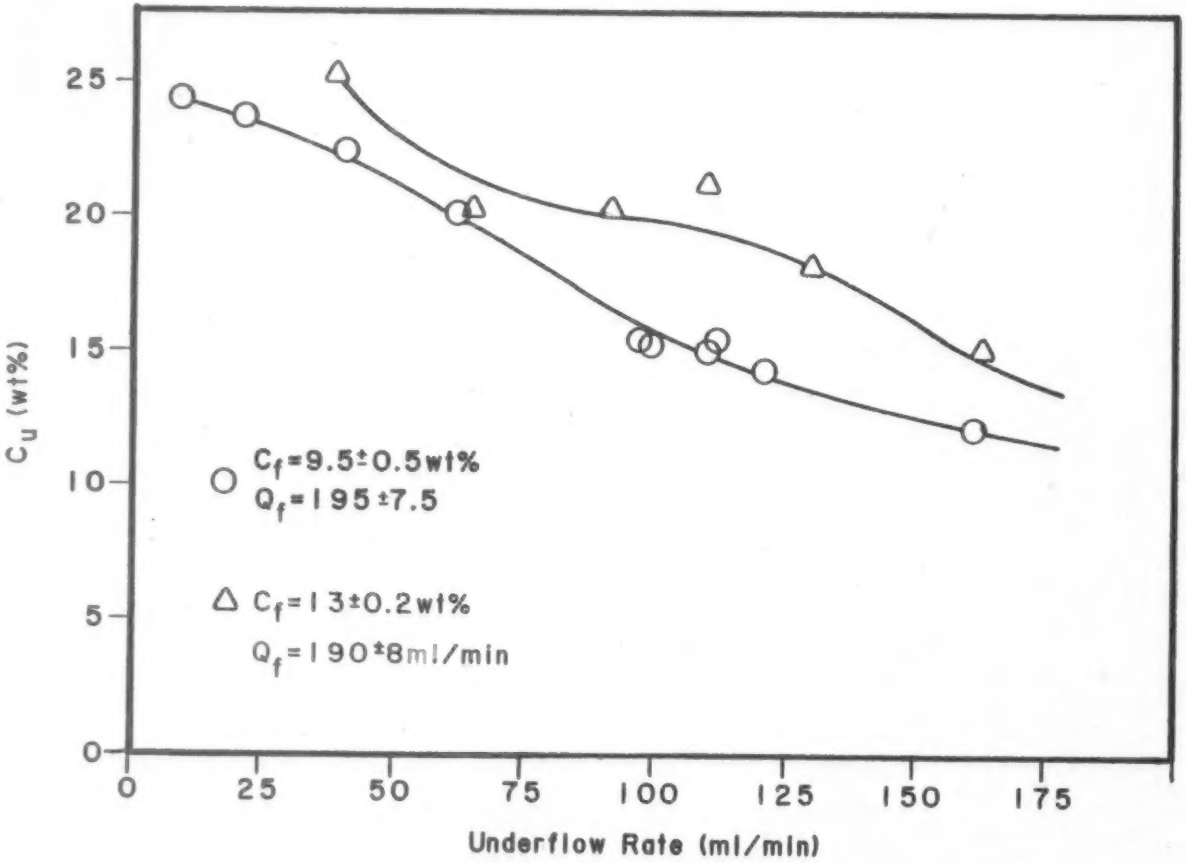


Figure 9.10. Spherical settler : effect of underflow rate on underflow concentration.

changes in Q_u, a decrease in Q_u causing a significant increase in C_u. Consequently, both the overflow concentration and the underflow concentration are very sensitive to the concentration of the feed in solids loading. However, the underflow rate cannot be too low; otherwise the accumulation will pursue in the thickener at a certain feed rate and eventually propagate into the clarifier. In such a case, the system would be out of capacity, and a clear overflow would not be attained. The optimum underflow rate should have the thickener at its maximum accumulation of solids; this accumulation should not propagate into the clarifier. Such a case depends on the separation area of the thickener, the capacity of the thickener (i.e., the residence time), the separation efficiency in the thickener, and the solids loading.

Section 10

EXPERIMENTAL EVALUATION FOR THE PILOT DEWATERING SYSTEM

Equipment Installation

A schematic of the pilot dewatering system is given in Figure 10.1. The experiments were conducted in a 6" ID 15' tall pilot column, where the section above the feed point was used as a splitter, the middle section between the feed point and the recycle stream output was used as a mixer, and the lower section below the output point of the recycle stream was used as a pilot thickener. The feed stock was made up in the feed tank at 10 wt% solids concentration. The overflow of the splitter fed into the clarifier was made up by a series of tube settlers (1.75" ID 6'-6" long) from the top section, where the feed port and overflow output were connected in series, but the underflow streams were combined together in parallel at the bottom. The number of tube settlers could be increased in order to match the increasing feed to the clarifier. For each tube settler, the feed port and underflow pick-up point were put at the lowest possible section of the cross-section, and the overflow outputs were put at the highest section possible. To avoid too much draw-off from the overflow of the clarifier because of unbalanced hydraulic forces, the collection point of the overflow (by the use of a funnel) was set at just the same level as the output point of the splitter. All the tube settlers were set at 30° to the horizon.

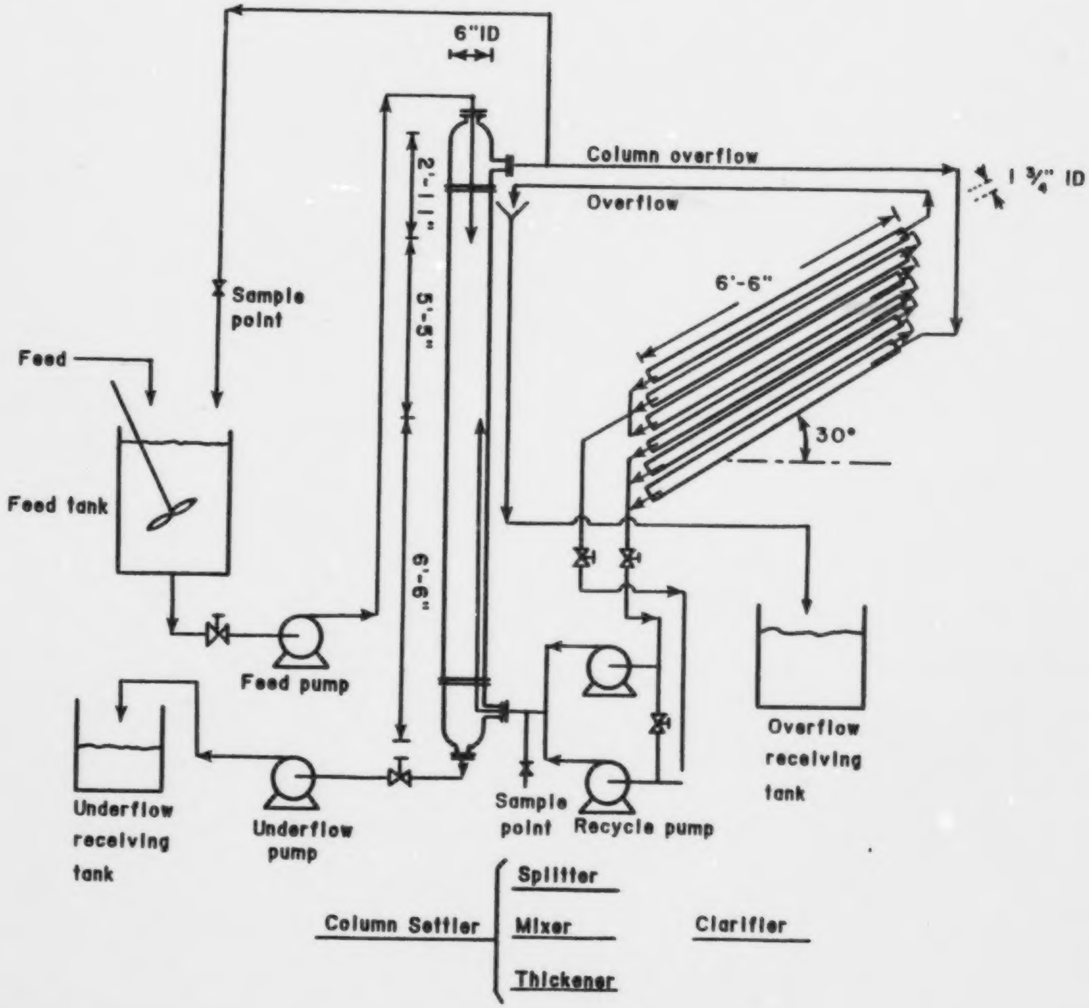

6"ID
Column overflow
1 3/4" ID
Overflow
6'-6"
2'-11"
5'-5"
6'-6"
Sample point
Feed
Feed tank
30°
Feed pump
Overflow receiving tank
Underflow receiving tank
Underflow pump
Sample point
Recycle pump
Column Settler
Splitter
Mixer
Thickener
Clarifier

Figure 10.1. A sketch of the clarifier/thickener system used in this work.

Sampling and Measurement

When the pilot system was operated at steady state (which could be roughly sensed by an examination of whether the interface was kept constant in the tube settler), and clear overflow was attained, then the feed rate and concentration, the recycle rate and concentration, the underflow rate and concentration, and the rate and concentration of the overflow from the splitter were measured. Also, the steady state evaluation and separation efficiency were examined according to the equations shown in Section 6.

The volumetric flow rates were attained by filling a graduated 1000 ml measuring cylinder for a certain time period.

General Method of Operation

During experimental runs, the pilot system was left running 24 hours a day, 7 days a week. The underflow pump was not used when the system was started, so that there could be an accumulation of solids in the thickener. Also, the settling tubes were used in series, in accordance with the solids accumulation occurring in the tubes, and the volumetric flow rate of the recycle pump was increased gradually to keep the interface in the clarifier, until it was constant in the middle of the top tube. In the meantime, the volumetric flow rate of underflow pump was adjusted according to the mass balance of the system. Usually it was set below the calculated rate before solids accumulation was completed. After the solids were packed to the top of the thickener (this point was manifested by the abrupt rise of the interface in the clarifier), the underflow pump was adjusted until a constant interface was obtained. In such cases, the system was at its steady state, and the underflow concentration from the thickener reached its maximum point at that feed rate and concentration.

The underflow concentration from the clarifier was affected by the recycle rate, as well as the underflow concentration of the thickener, i.e., up to a certain point, the higher the recycle rate, the greater the number of settling tubes needed, and the lower the underflow concentration from the clarifier. If the underflow rate from the thickener was set much less than the regular value which it should have, then solids would accumulate in the thickener. Eventually, solids would fill the thickener, and go over to the splitter and the clarifier. If this did happen, then the interface in the settling tube would rise, and ultimately a clear overflow could not be attained even if the recycle rate was increased to a maximum.

All in all, during operation, recycle rate should be increased or decreased according to which rate produces as clear an overflow and as large an underflow concentration from the thickener as possible. If, because of system upsets, etc., the overflow became too turbid, its clarity could be improved by simply decreasing the recycle rate. If the interface dropped to the lower tube, the recycle rate could be increased slightly to increase the underflow concentration from the thickener. During normal operations the recycle rate would be set at some mid-range, outside of which either the overflow would be cloudy or a lower underflow concentration from the thickener would result.

During operation, the wall effects on the settling tubes seemed to become pronounced with time and caused plugging of the settling tubes (the disadvantage of using tube settlers) resulting in a system upset.

To avoid such an upset, the periodic cleaning of the tube settlers one-by-one was necessary in this system.

Results

The pilot system was used for an evaluation of whether using the clarifier as proposed enhanced thickener performance and whether varying the recycle rate affected overall separation efficiency. First of all, as a determination of whether the clarifier did serve to enhance the thickener performance, the recycling of the clarifier underflow was discontinued (that is, the clarifier was bypassed) after the system had been run at steady state conditions. With the use of two different feed rates, 105 ml/min and 80 ml/min, the underflow concentrations decreased to steady-state values of about only 16% and 20% respectively, as shown in Figure 10.2. When the clarifier was reconnected, the data for the two feed rates (the recycle rate was set at about 30 ml/min) were collected after nearly steady-state conditions had been established. These are shown in Table 10.1 and 10.2. By comparing the results above, one can see that the clarifier apparently did serve to enhance the thickener performance significantly. The average underflow concentrations obtained were 37 ± 3% and 33 ± 3%. These were considerably less than the 35 to 45% obtained earlier with the larger test unit. This difference was assumed to be due to the action of wall effects, the much lower settling area of the thickener, "bridging" of the solids, and the lower compressive effect of solids overburden.

The downward solids flux, with and without the clarifier in line, was about 5.39 lbm solids/ft^2-hr and 4.33 lbm solids/ft^2-hr respectively

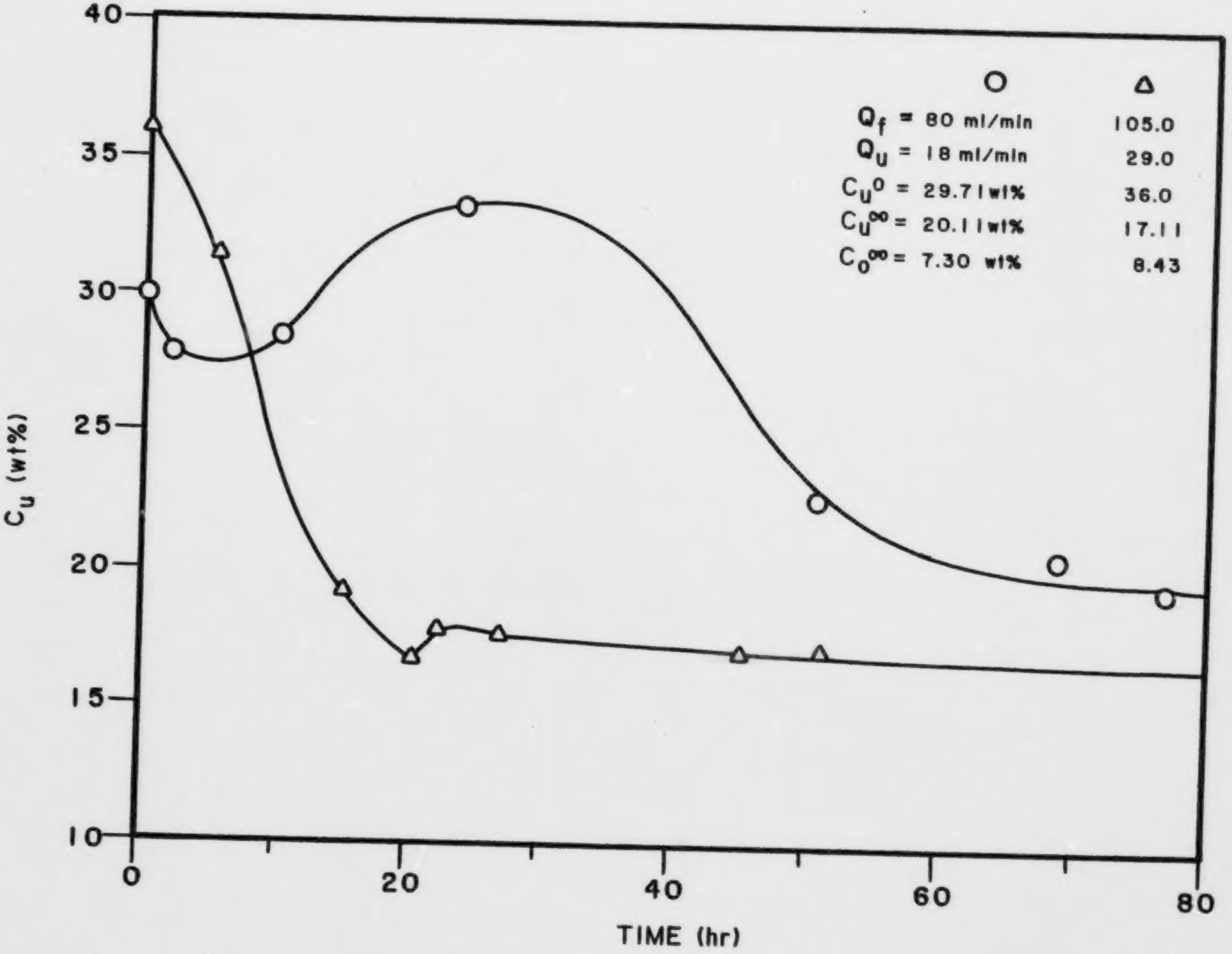


Figure 10.2. Effect of clarifier on the thickener for the pilot system.

Table 10.1 Operational Data of the Pilot System for Q_f = 80 ml/min

Flow rates, ml/min			Underflow Concentration, C_u	Clarification Efficiency
Feed Q_f	Overflow Q_o	Underflow Q_u	(wt% solids)	$(Q_f/0_f)(100\%)$
80	63	17	35.03	78.75
76	58	18	34.50	76.31
79	61	18	33.60	77.20
80	62	18	32.71	77.5

Feed concentraiton, C_f = 9.50 ± 0.30 wt% solids
Clear overflow from the system
Overflow from the column thickener to the clarifier = 8.75 wt%
Recycle rate = 30 ml/min

Table 10.2 Operational Data of the Pilot System for Q_f = 105 ml/min

Flow rates, ml/min			Underflow Concentration, C_u	Clarification Efficiency
Feed Q_f	Overflow Q_o	Underflow Q_u	(wt% solids)	$(Q_f/O_f)(100\%)$
102.8	86.0	16.8	41.0	83.7
109.6	92.5	17.1	35.3	84.3
106.6	89.0	17.6	34.8	83.5
105.4	88.0	17.4	35.7	83.5

Feed concentraiton, C_f = 9.50 ± 0.30 wt% solids
Clear overflow from the system
Overflow from the column thickener to the clarifier = 8.75 wt%
Recycle rate = 30 ml/min

(as calculated by equation 12 shown in Section 6) for a feed rate of 105 ml/min. According to the solids flux curve described earlier and an underflow concentration of 16% (the value obtained without the clarifier in line), the critical solids flux was 3.25 lbm solids/ft^2-hr (Figure 7.3). As stated earlier (Section 7) this critical flux should be a conservative value, and indeed, it is 25% lower than that obtained when the clarifier was not used. Also, since the overflow concentration from the thickening unit was very high (about 8 to 9 wt.%), more area would be required than that actually used; therefore, the critical value of solids flux computed by the use of Figure 7.3 is much too high. By the use of an underflow concentration of 36% (that obtained when the clarifier was in line) the critical flux was estimated to be 2.03 lbm solids/ft^2-hr, which is 60% lower than that initially obtained. In summary, the use of a clarifier as proposed does enhance thickener performance. However, the exact amount by which thickener area can be reduced or solids throughput can be increased cannot be determined from these small-scale pilot tests.

For an evaluation of the effect of recycle rate on system performance, the system was run with different recycle rates at two different feed rates. The results are shown in Tables 10.3 and 10.4. Figure 10.3 shows the results of the plotting of the underflow concentration (C_u) versus the dimensionless term (Q_{uc}/Q_f). As the clarifier is connected, (i.e., as the recycle stream is run), the underflow concentration increases abruptly after which it gradually increases at a certain slope until a maximum point is reached. Thereafter, the underflow concentration decreases gradually along with

Table 10.3 The Effect of Recycle Rate on the Underflow Concentration for the Pilot System Shown in Figure 10.1 at $Q_f = 77 \pm 3$ ml/min.

	Flow rates, ml/min		Underflow Concentration, C_u (wt% solids)		Clarification Efficiency
Recycle Q_R	Overflow Q_f	Underflow Q_u	C_u	C_{uc}	$(Q_f/O_f)(100\%)$
30	78	18	32.86	28.87	76.9
41	74	18	33.25	27.26	77.42
76	77.5	17.5	36.60	24.43	77.42
84	78.75	18	36.3	24.44	77.14
104	78.33	17	39.3	21.86	78.30
154	78.0	18	31.05	18.88	76.92
192	80.25	20	27.58	14.38	75.70

All above data are taken by the average of a series of experiments.
Feed concentration, $C_f = 9.50 \pm 1$ wt%.
Clear overflow from the system.

Table 10.4 The Effect of Recycle Rate on the Underflow Concentration for the Pilot System Shown in Figure 10.1 at Q_f = 100 ± 5 ml/min.

Flow rates, ml/min			Underflow Concentration, C_u (wt% solids)		Clarification Efficiency
Recycle Q_R	Overflow Q_f	Underflow Q_u	C_u	C_{uc}	$(Q_f/O_f)(100\%)$
33	100.25	17.50	32.90	-	82.54
57	98.00	16.33	33.30	-	83.17
100	101.33	16.50	33.70	-	83.72

All above data are taken by the average of a series of experiments.
Feed concentration, C_f = 9.61 ± 0.65 wt%.
Clear overflow from the system.

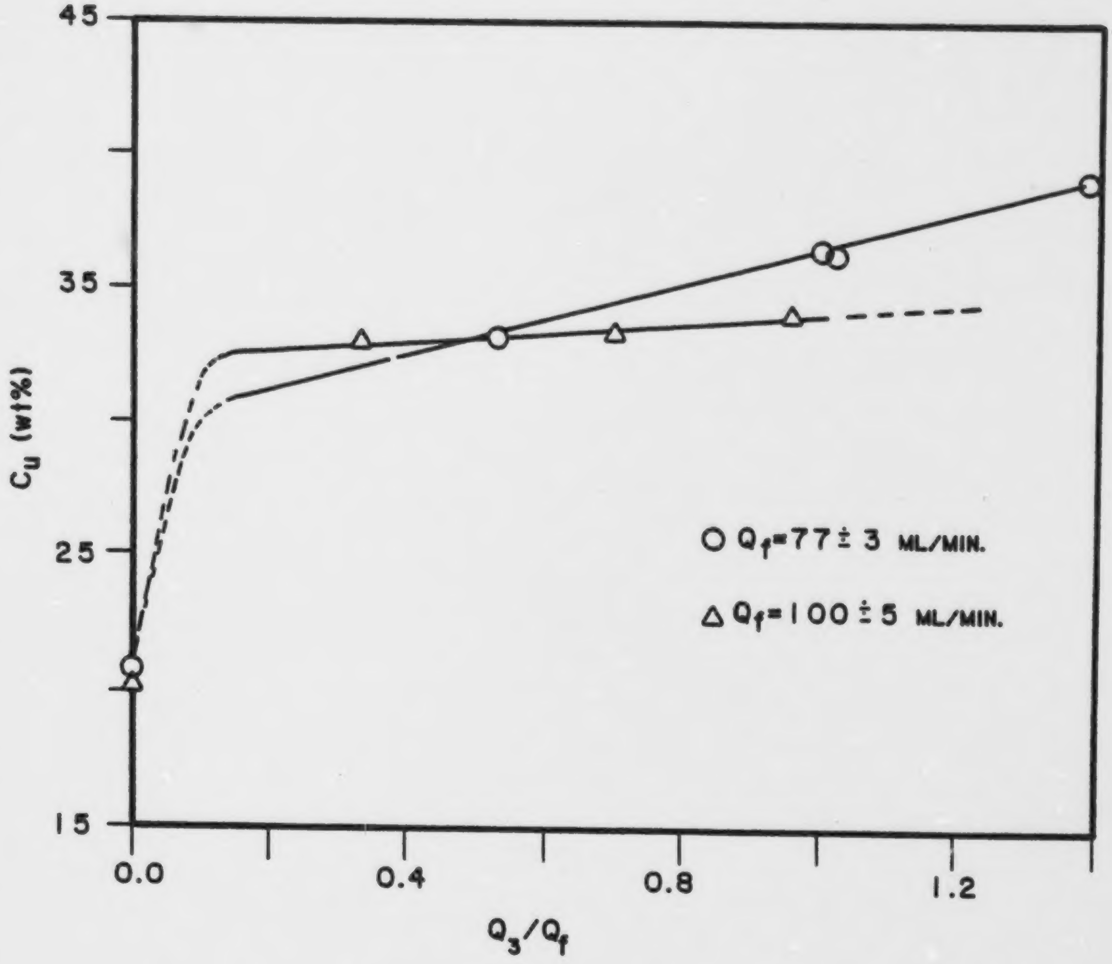


Figure 10.3. The effect of recycle rate on the underflow concentration at two different feed rates

the increase in recycle rate. If the recycle rate is infinite, as mentioned in Section 9 (equations 14, 15, 16, and 17) the functions of the mixer, the splitter and the clarifier appear to be of no use at all. Therefore, the underflow concentration approaches that obtained without the clarifier, i.e., 16% and 20%. In the meantime, the higher the recycle rate, the lower the slope of the line. For the feed rate fixed at 77 ml/min, the plot of recycle rate (Q_{uc}), and the concentration of recycle stream (C_{uc}) versus the underflow concentration (C_u) are shown in Figure 10.4. The higher the recycle rate, the lower the concentration of recycle stream which decreases linearly. As mentioned before in this section, the increase in recycle rate will increase the total solids residence time in the clarifier, and then increase the total function of the clarifier without the sacrifice of the decrease in the underflow concentration. The concentration difference between C_u and C_{uc} also gives the same results.

In summary, the separation efficiency of the proposed dewatering system is apparently sensitive to the changes in the rate of recycle from the clarifier to the thickener.

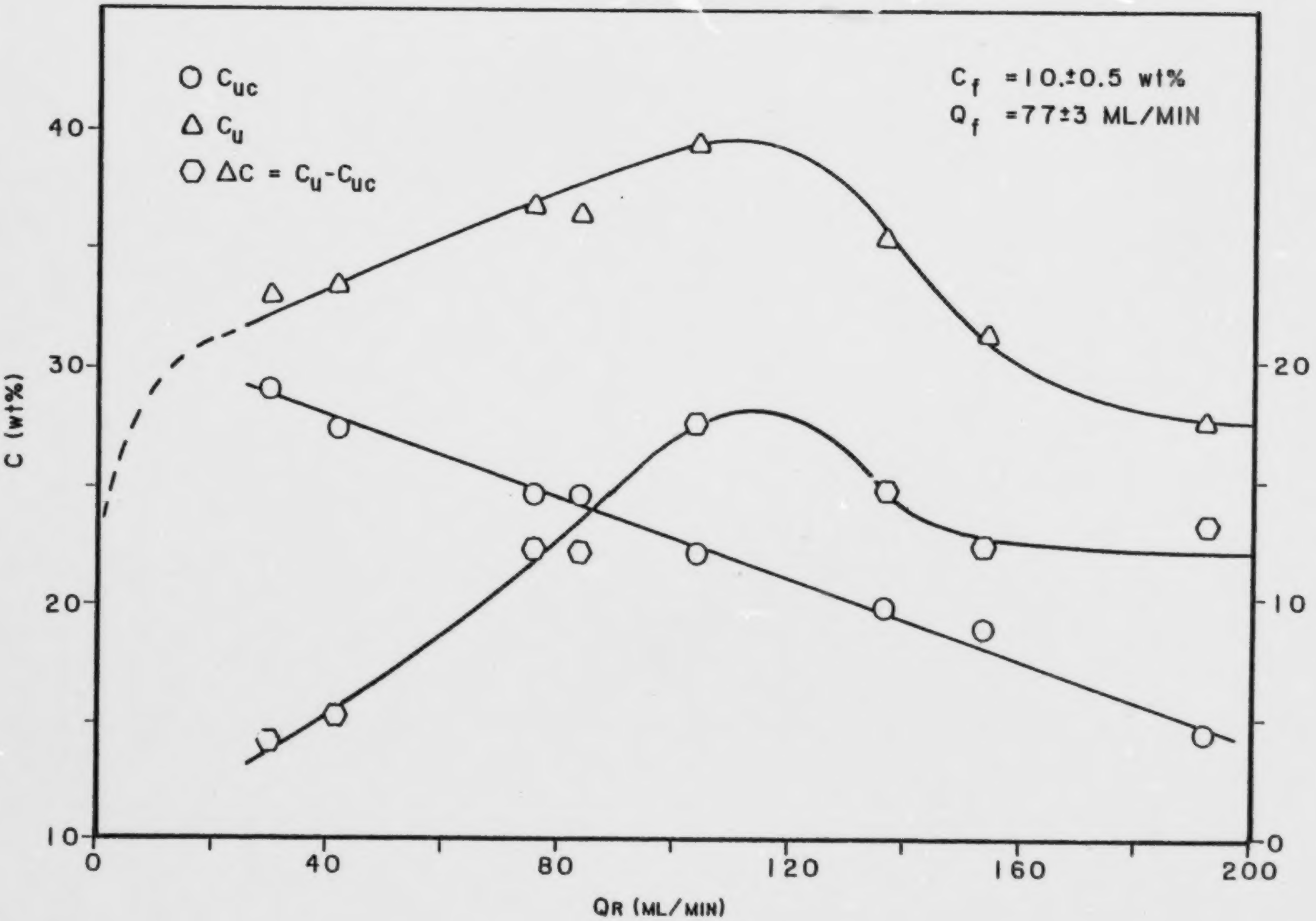


Figure 10.4. Recycle rate vs underflow concentration for a single feed rate (Q_f)

Section 11

PHASE II

CONCLUSIONS

An effective system for dewatering FGD wastes using gravity sedimentation has been developed in this work. The proposed system combines a laminar-type clarifier unit with a conventional thickener unit and introduces a recycle stream between them. The two units can not only be operated separately, but also allowed more flexibility and controllability.

In the study of the conventional thickener, it was found that the residence time of FGD waste solids in the compression zone of a gravity thickener has a very significant effect on underflow concentration and hence separation efficiency. For the FGD wastes tested, when the residence time of solids is sufficiently high, underflow concentrations higher than 40 to 45 wt% solids can be obtained. But, in order to get faster separation, especially for very low concentration feed, and to reduce the area requirements for the thickener, a clarifier is also needed.

The results of the small-scale pilot system tests have shown this system to be effective not only in obtaining the above requirements but also in raising the underflow concentration by 20 to 25 wt% solids with the use of a thickener alone up to the range > 35 to 40 wt% solids. If the separation area and the residence time are large enough (as the first unit used for the study of the conventional thickener) the system should yield an underflow concentration as high as 45 to 50 wt% solids and still produce a clear overflow at high throughput rates and with much lower area requirements than commercially available, conventional thickeners.

As shown in Figure 7.3, the higher the breakdown point of the floc structure, the higher the settling flux. But the actual point beyond K is still in question. More work can be directed toward determining the exact operating points for a proposed dewatering system. This breakdown point, however, depends on the separation characteristics of the slurry. For example, a high breakdown point might result from the vigorous beating of the concentrated slurry. These characteristics that will give higher breakdown points in the flux plot are presently being investigated.

Section 12

Sizing Studies and Economic Evaluation

PHASE III

FEASIBILITY STUDY: EVALUATING POTENTIAL SAVINGS WITH THE PROPOSED SYSTEM

The potential for cost savings through use of the proposed FGD waste dewatering system were evaluated by comparison with a conventional system. Two cases were considered as follows:

Case I: Replace the thickener in the conventional dewatering train (Fig. 12.1) with the proposed dewatering system (Fig. 12.2). The underflow concentration is to remain the same (~35% solids by weight).

Case II: Replace the thickener and the rotary drum vacuum filter in the conventional dewatering train with the proposed dewatering system (Fig. 12.3). Here the proposed system is to be designed to yield an underflow concentraion of ~55% to 60% solids by weight.

In Case I the primary advantage is that of substituting clarifier area for thickener area, clarifier area being assumed roughly one-third as expensive. To calculate the thickening area required for a certain underflow a plot of settling flux vs. solids concentration is used (Fig. 12.4). The following assumptions were used in the construction of the plot:

1) The settling rate is a function of concentration only.
2) One dimensional continuity is assumed.
3) Shear within and prior to entering the thickener is neglected.
4) The effect of compression is neglected.

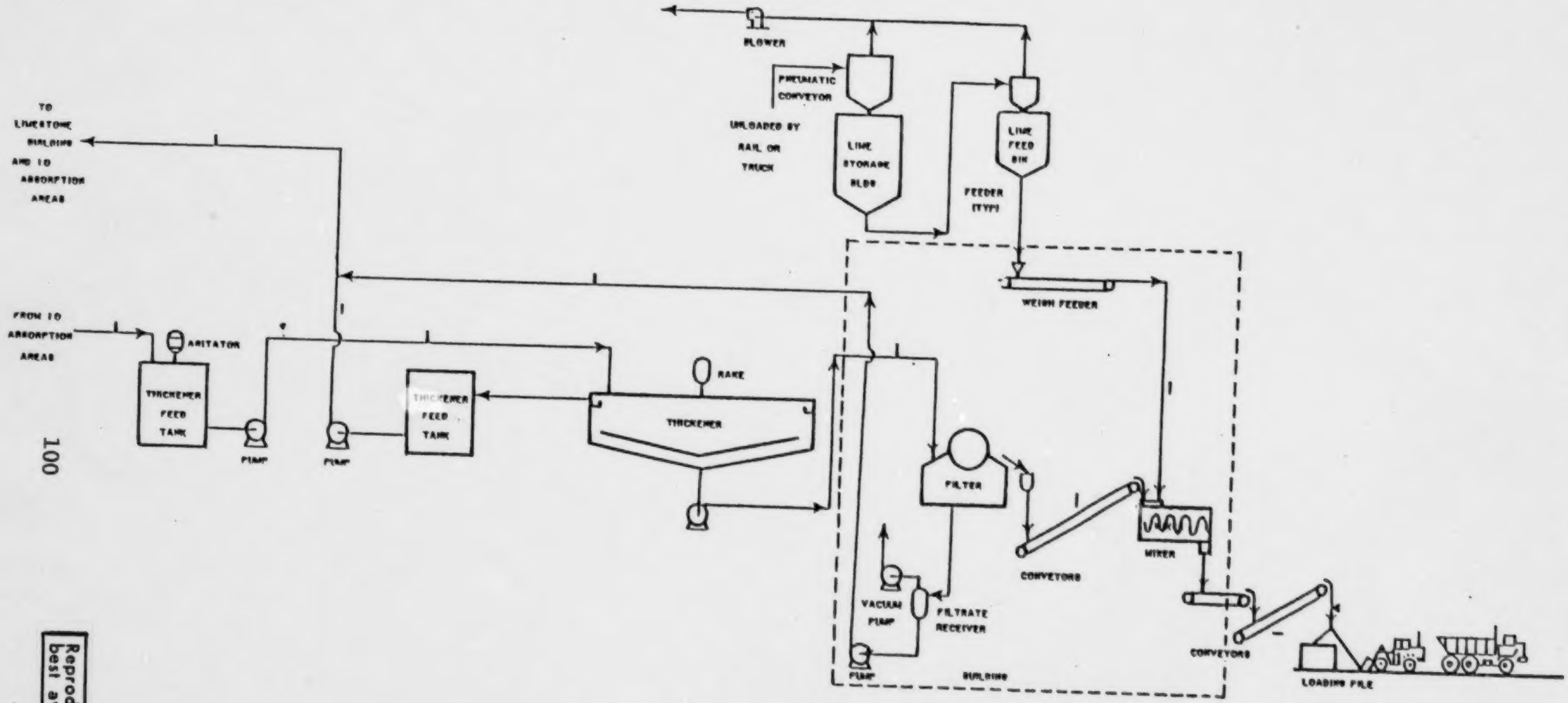

TO LIMESTONE BUILDING AND TO ABSORPTION AREAS
FROM TO ABSORPTION AREAS
AGITATOR
THICKENER FEED TANK
PUMP
PUMP
THICKENER FEED TANK
RAKE
THICKENER
BLOWER
PNEUMATIC CONVEYOR
UNLOADED BY RAIL OR TRUCK
LIME STORAGE BLDG
LIME FEED BIN
FEEDER (TYP)
WEIGH FEEDER
FILTER
VACUUM PUMP
FILTRATE RECEIVER
PUMP
BUILDING
CONVEYORS
MIXER
CONVEYORS
LOADING PILE

Figure 12.1. Conventional System

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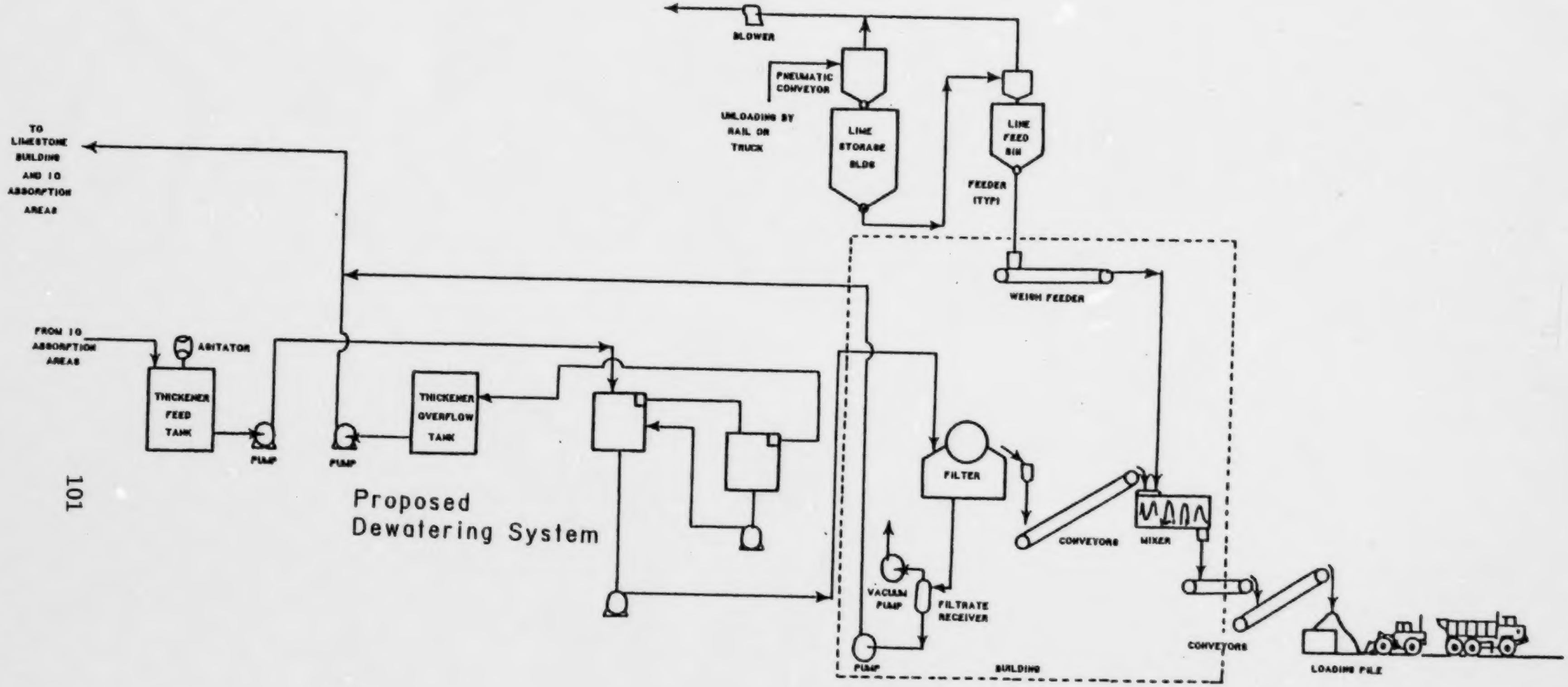
BLOWER
PNEUMATIC CONVEYOR
UNLOADING BY RAIL OR TRUCK
LIME STORAGE BLDG
LIME FEED BIN
FEEDER (TYP)
TO LIMESTONE BUILDING AND 10 ABSORPTION AREAS
WEIGH FEEDER
FROM 10 ABSORPTION AREAS
AGITATOR
THICKENER FEED TANK
PUMP
PUMP
THICKENER OVERFLOW TANK
Proposed Dewatering System
FILTER
CONVEYORS
MIXER
VACUUM PUMP
FILTRATE RECEIVER
PUMP
BUILDING
CONVEYORS
LOADING PILE

Figure 12.2. Case 1: Replacement of Thickener by Proposed Dewatering System

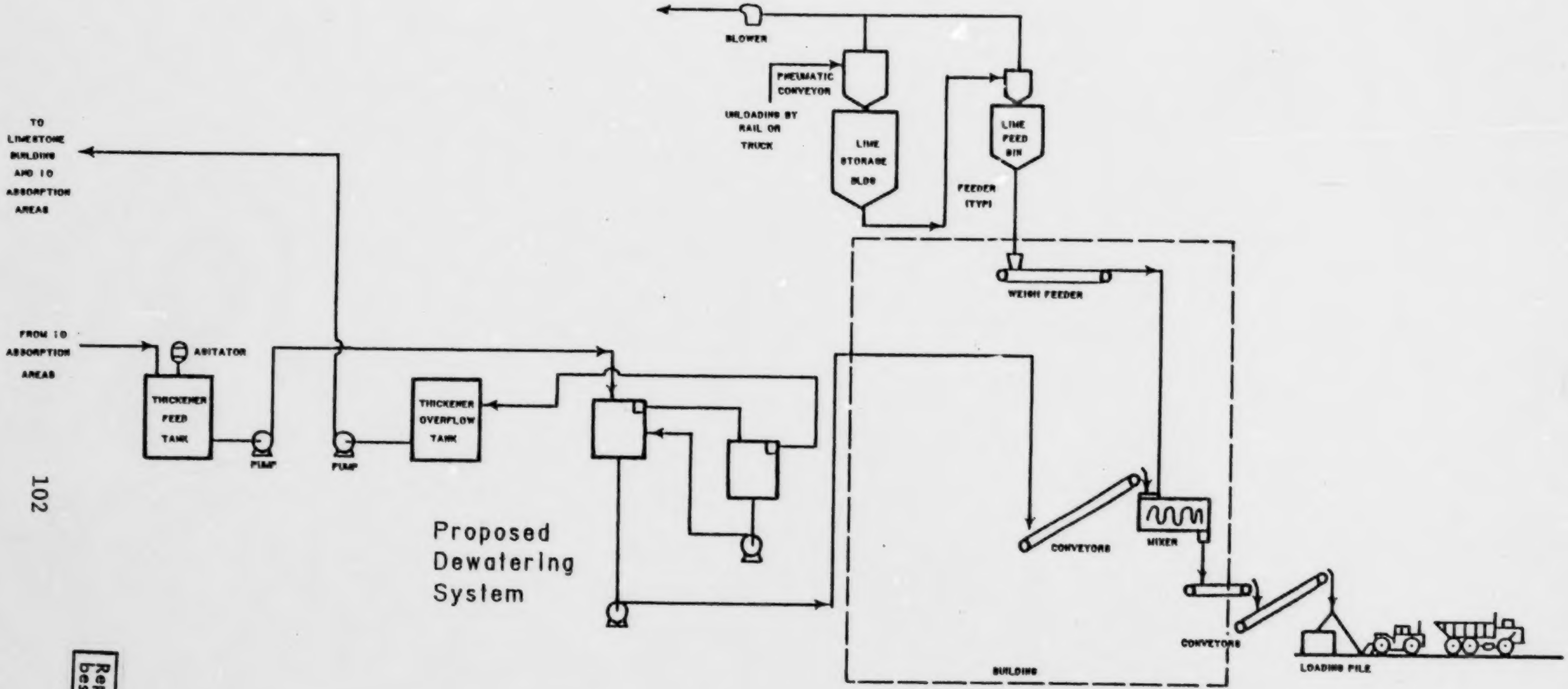
TO LIMESTONE BUILDING AND IO ABSORPTION AREAS
FROM IO ABSORPTION AREAS
AGITATOR
THICKENER FEED TANK
PUMP
PUMP
THICKENER OVERFLOW TANK
Proposed Dewatering System
BLOWER
PNEUMATIC CONVEYOR
UNLOADING BY RAIL OR TRUCK
LIME STORAGE BLDG
LIME FEED BIN
FEEDER (TYP)
WEIGH FEEDER
CONVEYORS
MIXER
CONVEYORS
BUILDING
LOADING PILE

Figure 12.3. Case 2: Replacement of Thickener and Filter by Proposed Dewatering System

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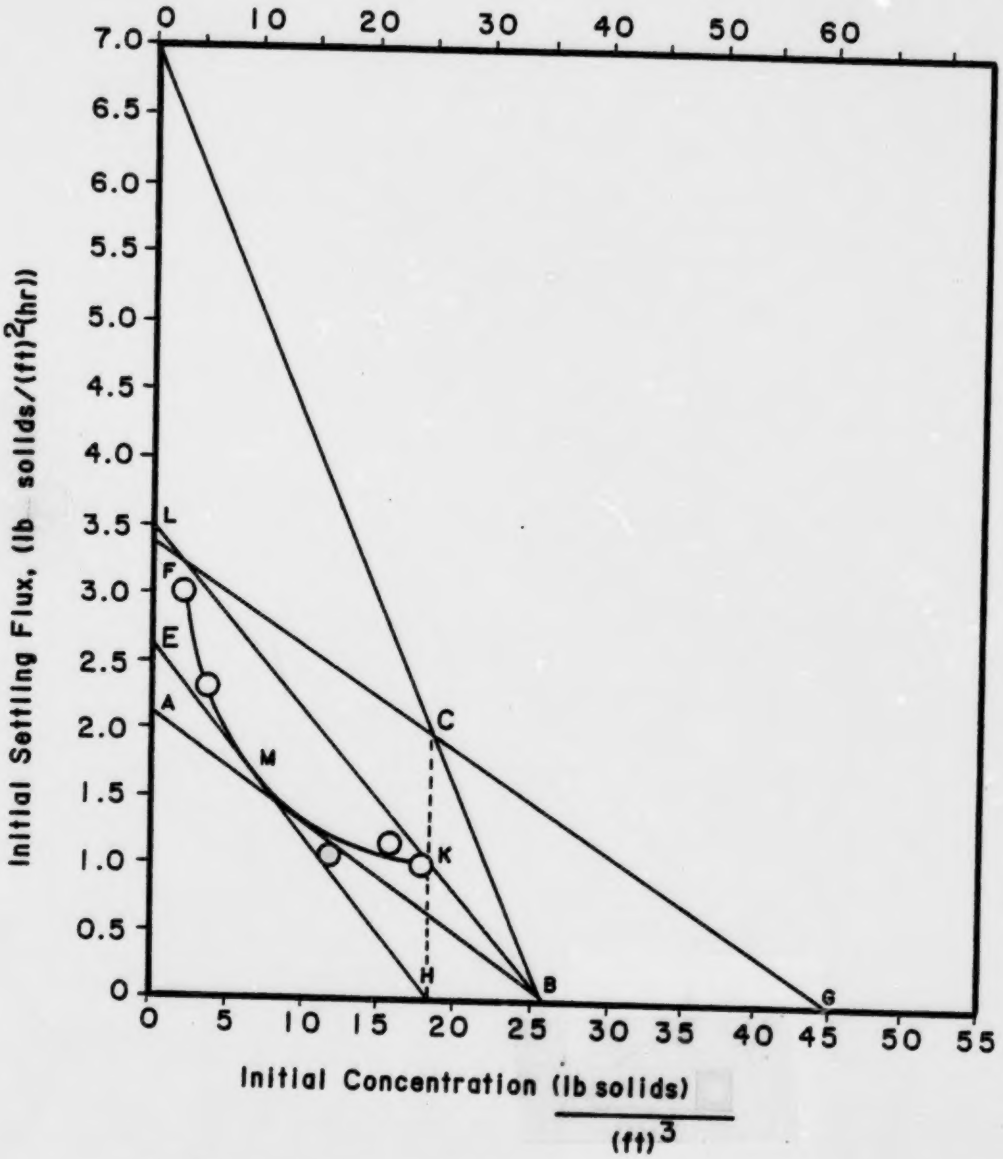

Figure 12.4. Flux Plot for the FGD Wastes of Louisville Gas and Electric

The above assumptions imply a linear relationship for the rate of upward propagation of concentrated layers of solids. That is, as settled material accumulates in a batch settling system it propagates linearly upward toward the solid/liquid interface. A linearization in effect is thus used, which is a common engineering practice for analyzing complex (non-linear) systems.

For a conventional thickener, the required area is estimated by drawing a line passing through the desired underflow concentration and tangent to the flux plot. The reading on the vertical axis gives the settling flux from which the required area can be determined.

From line AB in Figure 12.4 it is estimated that about 0.44 ft^2 hr/lb_m solids is required for a conventional thickener to obtain an underflow concentration of 35 wt.%. The calculations of the area required for the proposed system are based on two primary assumptions:

1) The clarifier will pre-concentrate the slurry to about 25% by weight.
2) The breakdown of the flox structure between the clarifier and the thickener will cause the solids flux in the bottom of the thickener to increase to roughly twice that normally possible (point C in Fig. 12.4).

A line drawn through point C and the desired underflow concentration will then give the area required for the thickener. From Line DCB the thickener area for the proposed system is estimated to be 0.14 ft^2-hr/lb_m solids. Assuming the cost of clarifier area to be roughly one-third that of thickener area, the total area required on a comparative basis, is 0.33 (0.37 ft^2 hr/lb) = 0.12 ft^2 - hr/lb.

∴ Area Normally Required = 0.44 ft^2-hr/lb solids

Thickener Area Required for
Proposed System = 0.14 ft^2-hr/lb solids

Clarifier Area Required for
Proposed System = 0.12 ft^2-hr/lb solids (on a comparative basis assuming clarifier area to be about 1/3 as expensive as thickener area).

Total Area Required for
Proposed System = 0.14 + 0.12 = 0.26 ft^2-hr/lb

Total Reduction in
Area Requirements = $\frac{0.44 - 0.26}{0.44} \times 100 = 41\%$

Thus the total area required, for the proposed dewatering system, is approximately 59% of the area required for the conventional thickener. this represents substantial reductions in capital costs which are a strong function of area required.

The potential gains in Case I do appear to be significant. However, it is in Case II that the greatest economic gains are expected to be realized. Following similar procedures as in Case I and using line FG in Figure 12.4, the thickener area is estimated to be about 0.33 ft^2-hr/lb (for an underflow concentration of 60% by weight). Therefore,

Area Normally Required = 0.44 ft^2-hr/lb

Thickener Area Required
for Proposed System = 0.33 ft^2-hr/lb

Clarifier Area Required
for Proposed System = 0.12 ft^2-hr/lb (on a comparative basis assuming clarifier area to be about 1/3 as expensive as thickener area)

Total Area Required
for Proposed System = 0.33 + 0.12 = 0.45 ft^2-hr/lb

In other words, using about the same area as that required for the conventional thickener, the slurry can be concentrated, using the proposed system, to about 60 wt.%. This will potentially allow elimination of the filtration train. Elimination or reduction of the filtration stage would obviously result in substantial savings.

To better illustrate the above estimations, assume the conventional thickener area is 49,087 ft^2 that is, the thickener has a diameter of 250 ft. To concentrate the slurry to 35%, the estimated size requirements of the proposed sytem would be:

$$\text{Thickener Area} = (49{,}087\ ft^2)\left(\frac{0.14}{0.44}\right) = 15{,}619\ ft^2$$

i.e. a 141 ft. diameter thickener would be required as opposed to 250 ft.

$$\text{Clarifier Area} = (49{,}087\ ft^2)\left(\frac{0.37}{0.44}\right) = 41{,}278\ ft^2$$

$$\begin{aligned}\text{Area Cost Equivalent of proposed system} &= 15{,}619\ ft^2 + 1/3(41{,}278\ ft^2)\\ &= 29{,}378\ ft^2 \text{ as opposed to } 49{,}087\ ft^2\end{aligned}$$

To obtain a 60% by wt. underflow,

$$\text{Thickener Area} = (49{,}087\ ft^2)\left(\frac{0.33}{0.44}\right) = 36{,}815\ ft^2$$

$$\text{Clarifier Area} = (49{,}087\ ft^2)\left(\frac{0.37}{0.44}\right) = 41{,}278\ ft^2$$

For this case, the filters conventionally required can be potentially eliminated. In summary then, the potential economic savings offered by the proposed dewatering system are significant.

A detailed comparative economic analysis for the proposed system was performed by the Tennessee Valley Authority (TVA) at EPA's request. In order to determine the potential savings, the conventional FGD waste handling and disposal system shown in Figure 12.1 (base case) was compared with the systems shown in Figs 12.2 and 12.3 (Case I and Case II, respectively). Conversion Systems, Inc. (CSI)* process was chosen as the base case system; this process was evaluated and discussed in the report entitled _Economics of Disposal of Lime-Limestone Scrubbing Wastes: Untreated and Chemically Treated Wastes_ (27). The table below summarizes the capital investment and annual revenue requirement estimates.

	Capital Investment $1000	Annual Revenue Requirements, $1000
Conventional-Base Case (IUCS)	10,717	5,291
Auburn- Case I (Proposed)	9,669	5,131
Auburn-Case II (Proposed)	9,713	4,970

The capital savings for the Auburn dewatering system (Case I) over the conventional dewatering system is approximately $1.05M. This is approximately 0.44% of the total capital investment for a limestone scrubber process and would result in Annual Revenue Requirement savings of approximately $160,000. Case I is based on a thickener underflow concentration (which is yet to be demonstrated) of 80% solids by weight. Case II is based on achieving a thickener underflow concentration of 65 weight % solids followed by filtration of the slurry to produce a gypsum filter cake with a concentration of 80% solids by weight. Even with Case II a capital savings of approximately \$1.0M (0.42% of the total capital investment of a limestone scrubber process) is still possible. Annual revenue requirement savings of approximately \$320,000 are projected for Case II.

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* formerly known as IU Conversion System, Inc. (IUCS)

Details of the above study are shown in tabular form in Appendix C.

Section 13

PRELIMINARY DESIGN WORK FOR LARGE SIZE PILOT SYSTEM

Many problems were encountered while performing the lab-scale development work of the dewatering system. Plugging of valves, pumps, and flow lines, etc., interrupted the work often and made it difficult to carry out steady-state experiments. The supply of FGD waste needed to perform the experiments was limited, and the waste had to be recycled while performing continuous studies. Excessive recycling of most sludges affects their settling behavior. The exact effect that recycling of the FGD waste had on the separation efficiencies of the lab-scale systems and on the settling rates, floc structure of the sludge, etc., cannot be predicted. In other words, the data collected using lab-scale set-ups cannot be used confidently for scale-up to a commercial system for many reasons such as these given here. Larger pilot scale experimentation had to be done before the dewatering system would be accepted for commercial-scale use.

A joint grant was received from the EPA and TVA for the planning, installation, and operation of a pilot FGD waste dewatering unit at the TVA Shawnee facility at Paducah, Kentucky. The objective of this research was to evaluate the performance of the dewatering system on a larger pilot scale, on-site at Shawnee. This evaluation had a three-fold purpose: to confirm and better establish the results of the lab-scale tests performed at Auburn, to provide information for scale-up purposes, and to provide operational data needed to better estimate the economic advantages offered by the dewatering system.

A flow diagram of the gravity sedimentation system proposed for the dewatering of FGD wastes is given in Figure 13.1. A classical type thickener and a laminar-type clarifier are used in combination. The role of the clarifier was to prethicken the FGD wastes as well as to clarify the overflow. The role of the thickener was to thicken the wastes as much as possible. The use of a laminar-type clarifier and a thickener is by no means a novel concept in thickener technology. In conventional systems, however, the clarifier is usually mounted above the thickener, and the thickener is usually just a relatively small sump tank. In the proposed system the thickener area can be as large or as small as desired, and the recycle stream can be fed into the thickener at any desired height. In other words, in the proposed system, the clarification and the thickening function of the system are decoupled as much as physically possible. in addition, unlike conventional systems, the recycle rate can be controlled. This permits a fine operational control of the dewatering system that is very valuable.

The feedport is located near the top rather than the bottom of the clarifier, as is the usual practice, and is expected to have a significant beneficial effect on its performance. It is interesting that, in most recently released product bulletins of Parkson's lamella thickener, the feed is no longer introduced at the bottom of the settling plates but at some higher point in the unit.

To prevent or at least minimize any disruption of operation while testing the proposed system, the clarifier overflow was to be fed to the

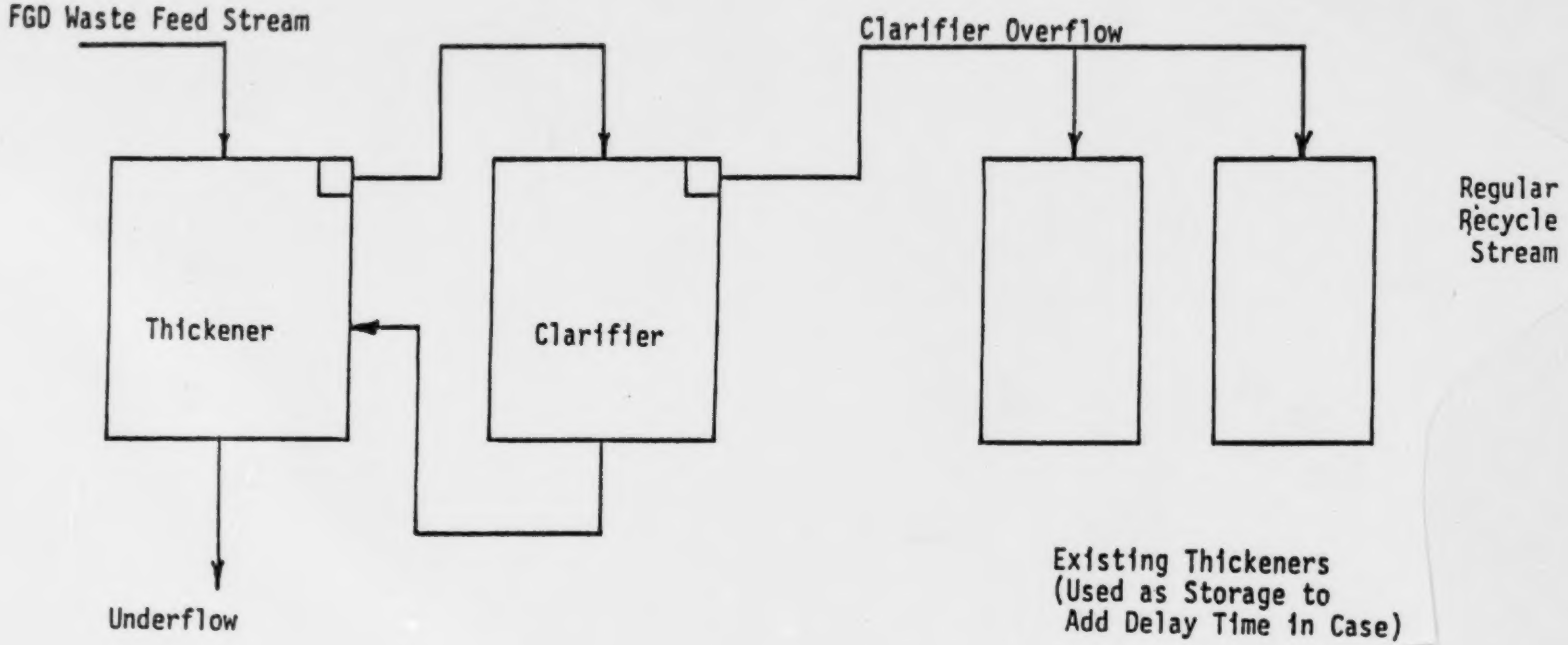

FGD Waste Feed Stream
Thickener
Underflow
Clarifier
Clarifier Overflow
Regular Recycle Stream
Existing Thickeners (Used as Storage to Add Delay Time in Case)

Figure 13.1. Proposed Flow Scheme Using Existing Thickeners In Series with Proposed Pilot Unit to Minimize Disruption of Overall Operation

existing thickeners while testing the proposed system. Then, if the proposed unit plugged, there would be sufficient storage capacity available to transfer smoothly to the normal mode of operation. By continuing to feed to the existing thickeners, the clarifier water in these existing thickeners would maintain the same concentration, etc., of recycle fluid. With this flow arrangement the potential for upsetting other ongoing tests, etc., would almost be completely eliminated. Figures 13.2 and 13.3 are the equipment drawings for the clarifier and the thickener respectively. Since the copies were made from blue-prints of the equipment it was not possible to make better reproductions of these drawings.

Figure 13.2. Equipment Drawing for the Clarifier

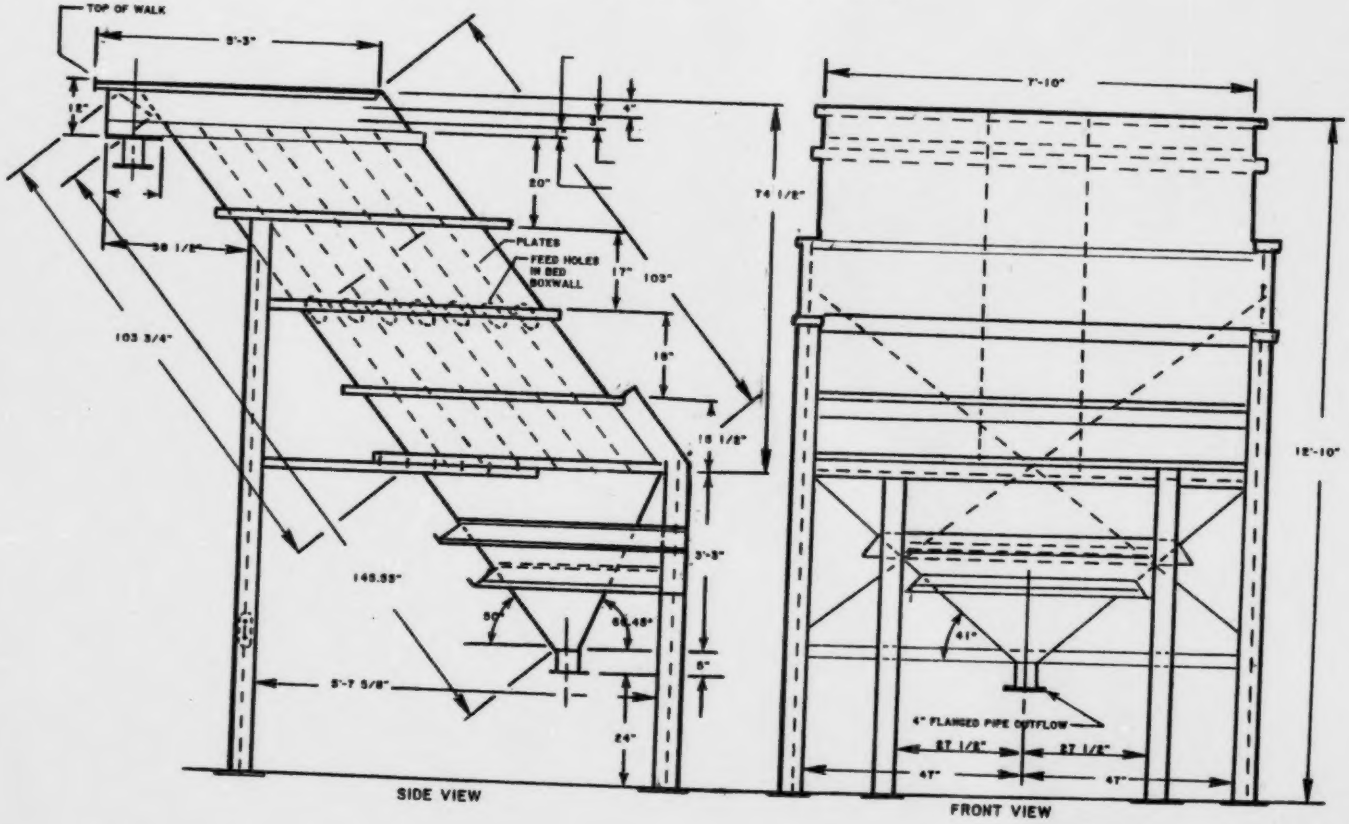
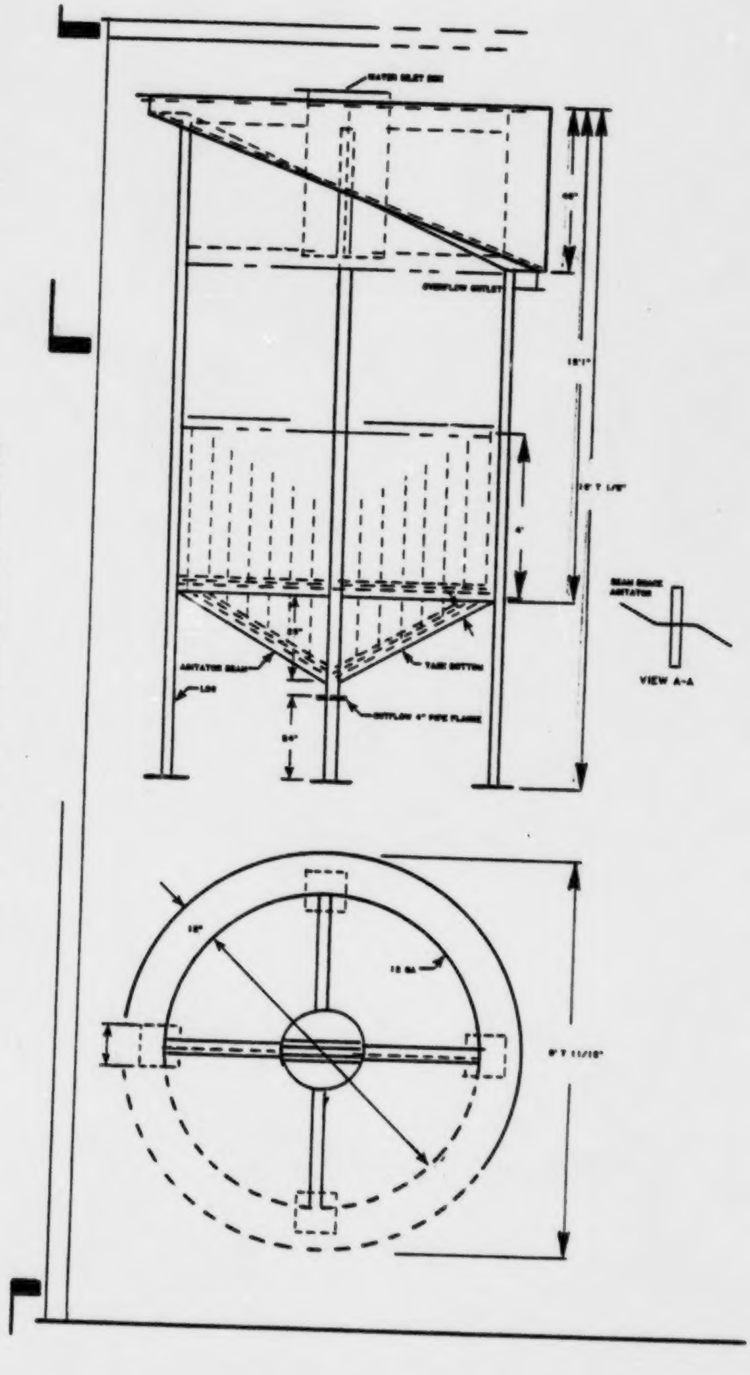

TOP OF WALK
5'-3"
12"
4"
3"
20"
74 1/2"
38 1/2"
PLATES
FEED HOLES IN BED BOXWALL
17"
103"
103 3/4"
18"
15 1/2"
3'-3"
145.55"
50°
63.45°
6"
5'-7 5/8"
24"
SIDE VIEW
7'-10"
12'-10"
41°
4" FLANGED PIPE OUTFLOW
27 1/2"
27 1/2"
47"
47"
FRONT VIEW

Figure 13.3. Equipment Drawing for the Thickener


WATER INLET BOX
60"
OVERFLOW OUTLET
15'1"
10' 7 1/8"
4'
BEAM BRACE AGITATOR
VIEW A-A
AGITATOR BEAM
TANK BOTTOM
LEG
OUTFLOW 4" PIPE FLANGE
24"
12"
10 GA
9' 7 11/16"

Section 14

EVALUATION OF EFFECTS OF PRE-TREATMENT OF FGD WASTE ON SETTLING BEHAVIOR RELATING TO SELECTION OF TYPE OF RECYCLE PUMP

The presence of calcium sulfite in FGD wastes often makes the wastes difficult to dewater and causes the wastes to exhibit thixotropic properties (i.e. an apparently solid FGD waste will reliquefy when agitated).

Based on the thixotropic properties of the waste sludge, it was hypothesized that the addition of shear to the recycle stream from the clarifier to the thickener would improve its settling characteristics. This is because the floc structure would be broken up to form smaller flocs of particles, causing a reduction in the extent of structural formation within the slurry and allowing additional amounts of water to be squeezed out. Tests were conducted on both a batch and a continuous scale to determine the effect of vigorous mixing on the settling behavior of FGD waste. The final outcome of these tests was to influence the choice of the type of pump to be used for the underflow from the laminar clarifier, as shown in Figure 14.1.

Batch Tests: FGD waste samples were taken with solids concentrations ranging from 2-57%. An experimental sample and a control sample were taken for each concentration. The experimental sample was vigorously stirred in a blender while the control sample was not. Both samples were then allowed to settle simultaneously in small plastic cylinders. The interface height was periodically measured for both samples. Results are tabulated in Table 14.1. From these results it can be seen that in batch tests vigorous mixing and addition of shear not only increased the settling rate but also resulted in a greater final concentration, i.e. at a given

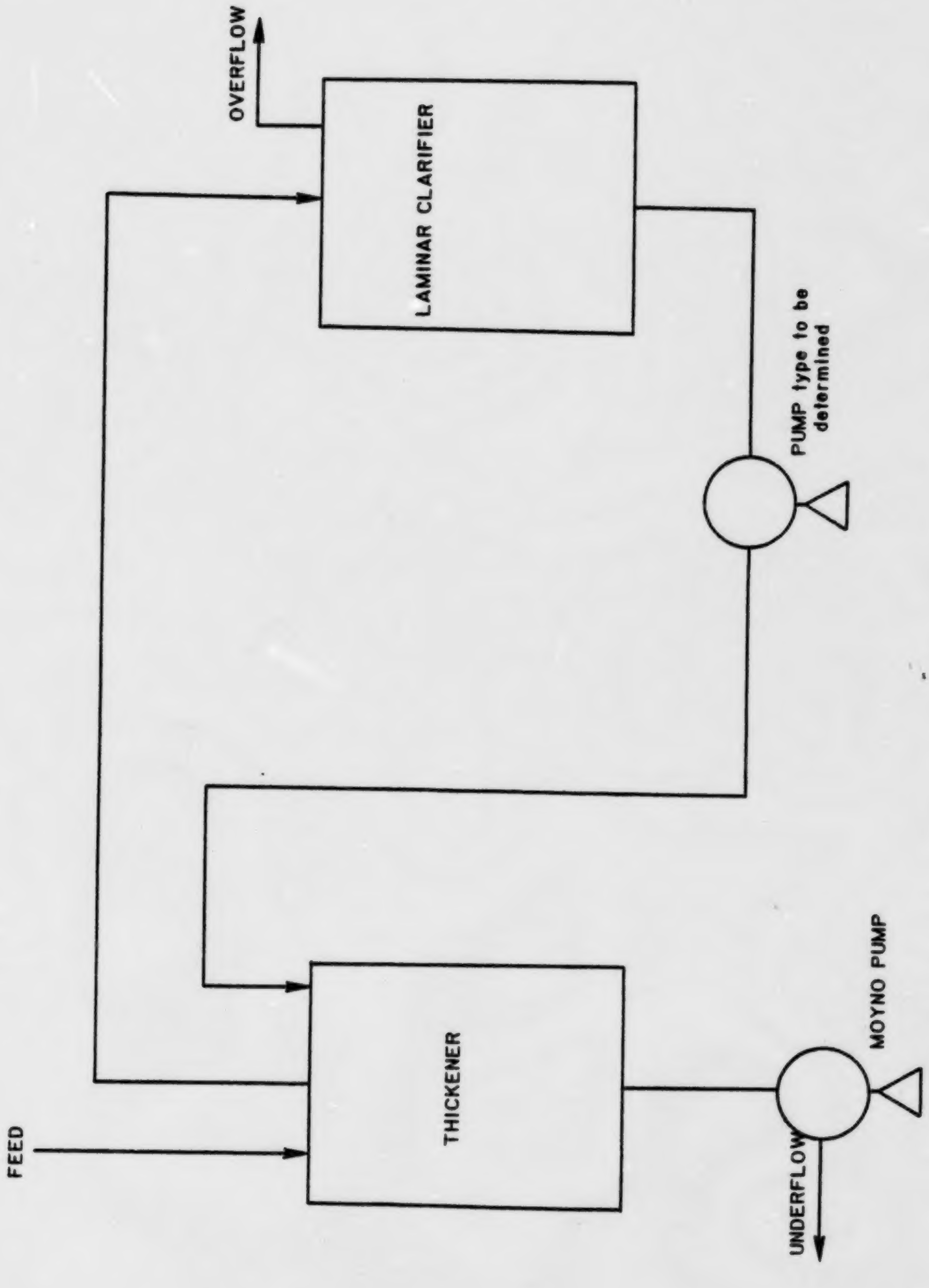

OVERFLOW
LAMINAR CLARIFIER
PUMP type to be determined
FEED
THICKENER
MOYNO PUMP
UNDERFLOW

Figure 14.1. Recycle pump type to depend on effect of Shear on settling behavior.

Table 14.1

Effect of Stirring on Settling Behavior of FGD Wastes (Batch Tests)

Initial Solids Concentration (wt.%)	Initial Interface Height (cm)	Final (steady state) Interface Height (cm)	
		Vigorously Stirred	Unstirred
2	1000	115	245
20	610	511	541
27	1000	641	664
30	1000	624	668
32	1000	775	822
36	900	719	758
42	1000	814	834
57	1000	835	850

point in time the experimental sample showed a lower interface height and also after a sufficiently long time, when the interface height became constant for both samples, the experimental sample had settled to a greater degree (lower interface height or higher final concentration). The results of the tests described above are shown in Appendix D.

Continuous Tests: A constant concentration feed was pumped at a constant rate into a clarifier, with the clear water from both the clarifier and the thickener being pumped back into the feed tank, while the concentrated slurry was pumped into a thickener (control case) and into a stirred tank (experimental case). The underflow from the thickener was recycled back into the feed tank thus forming a closed circuit system. The operating parameters of the system for both cases were the underflow rate and the constant interface between clear liquid and the slurry (once the system achieved steady state). The volumetric flow rates of the clear overflow and the concentration of the underflow were measured and compared for the control and experimental cases.

It is seen that although the experimental sample does show somewhat more desirable settling characteristics than the control sample, certain parameters for the samples were not the same. For example, the average feed concentration is higher while the feed rate is lower for the experimental sample.

Also, due to constant variations in operating conditions such as temperature, feed concentration, feed rate, the breakdown of pumps, and the fact that the system was frequently upset by the plugging of tubes, the continuous tests did not show as pronounced an effect of shear on settling behavior as did the batch tests.

Further results of both the batch and continuous tests are provided in Appendix D.

Section 15

SHAKEDOWN PILOT PLANT STUDIES

The pilot FGD waste dewatering unit was designed and constructed from October 1978 to February 1979. However, because of difficulties in obtaining TVA and EPA agreement on the tests to be conducted at Shawnee, delivery was delayed for several months.

The unit was delivered to the Shawnee site on September 26, 1979. It was installed and ready for operation by October 15, 1979. Because of a boiler breakdown, however, start-up of the unit was delayed. Also, TVA had some concerns about the unit and recommended a number of small modifications to facilitate maintenance and shutdown of the unit in case of pump failures or plugging of the lines. Some of the required modifications, e.g. the installation of a side platform to provide safe access to the top of the thickener, were necessary for compliance with TVA's safety policies at the Shawnee facility.

At a Shawnee Steering Committee meeting at Paducah on November 6, 1979, it was agreed to proceed with a shakedown operation of the unit: the unit was to be operated as long as weather permitted or until some equipment failure was experienced. The operational data collected during this shakedown period were to be used to design test runs which were to be made the following Spring, as soon as the weather permitted sustained operation of the unit. Any equipment failures experienced during the shakedown period were to be corrected during the interim downtime.

Once approval was attained from the steering committee, shakedown testing of the unit was started immediately. Valves were installed in the underflow lines for both the thickener and the clarifier. Several sample ports were installed to make sampling easier. The Moyno pumps were calibrated. A check valve was installed at the exit of the centrifugal pump for the sump tank. A valve was installed for isolating the clarifier from the thickener.

During the shakedown testing, the thickener was charged with slurry having a solids content of about 35 wt.%. Approximately 30% of the volume of the thickener was charged with slurry. This was done so that the system would attain steady-state operation as quickly as possible. The shakedown tests were carried out only during day shifts over a period of eighty hours (two weeks). During the other two shifts, the thickener underflow pump was kept on for recycling purposes. The underflow from the laminar clarifier was recycled back to the thickener.

The limited data collected during the shakedown testing were very encouraging in that underflow concentrations as high as 60% were obtained. Whether the unit could yield this high an underflow concentration, with sustained operation, had to be verified in a follow-up study.

The laminar clarifier was designed to handle a feed concentration of only 10% solids. The feed concentration was actually 16%. As a result, the system was not capable of delivering the desired clarification. This was the type of information that was expected to be gained from the shakedown testing. The clarification capacity of the system could be improved simply by adding more plates to the unit. Equipment funds were requested as part of a continuation of the pilot evaluation to allow more plates to be added to the clarifier, and eliminate the clarification problems experienced during the shakedown testing.

During the shakedown testing the liquid level controller in the sump tank did not work satisfactorily. A new control device was designed and installed. It performed well.

SECTION 16

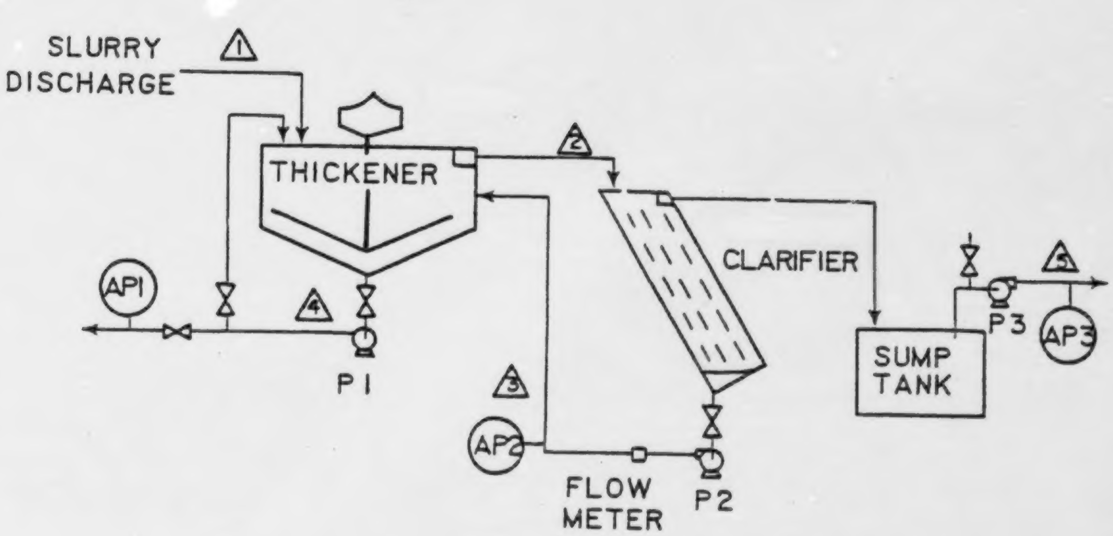
FINAL PILOT PLANT STUIES

A flow diagram of the unit used in the final testing, is shown in Figure 16.1. Start-up procedures and the original test plan are given in Appendix E. The test plan was developed in cooperation with TVA staff. Before attempting to schedule the tests, TVA decided to perform another analysis of the potential benefits offered by the proposed dewatering system (refer to Section 12 for the previous analysis). Upon completion of this analysis, it was decided to proceed with testing the proposed dewatering system.

Initially, it was hoped that testing of the dewatering system could be conducted in conjunction with other FGD system tests which TVA was conducting. However, TVA was concerned that operation of the dewatering system would adversely affect other FGD system objectives. For this reason, the time available for testing the dewatering system was limited (since the entire FGD system would have to be operated solely for the purpose of testing the dewatering system, making this test expensive). Scheduling the dewatering system testing was also difficult, because most of the time the FGD system was configured to generate wastes with high calcium sulfate levels (via forced oxidation), while it was desired that the dewatering system be tested with wastes with high calcium sulfite levels.

Both of these factors led to a number of delays in testing. However, a four-day, continuous (24-hour/day) test was conducted during the Fall of 1982. The following is a detailed discussion of the findings made during this testing.

Figure 16.1. Schematic of Auburn Dewatering Unit


SLURRY
DISCHARGE
1
THICKENER
2
CLARIFIER
5
AP1
4
P1
P3
AP3
SUMP
TANK
3
AP2
FLOW
METER
P2

STREAM NO.	1	2	3	4	5
DESCRIPTION	Slurry Discharge	Thickener Overflow	Clarifier Underflow	Thickener Underflow	Clarifier Overflow
Rate, lb/hr.	2186	3923	2332	595	1591
RATE, gpm	4.00	7.18	4.00	0.82	2.18
% SOLIDS	15.0	14.9	25.0	55.0	——
SPEC. GRAVITY	1.093	1.092	1.165	1.451	1.000

Based on $\rho_s = 2.3$ gm ml^{-1}, $\rho_l = 1.0$ gm ml^{-1}

The proposed system design was observed to allow the operational flexibility needed to decouple, for all practical purposes, the thickener from the clarifier. That is, the underflow rate of the thickener could be set very low without causing any clarification problems. Based on the findings of the shakedown studies, additional settling plates were added to the clarifier. The overall dimensions of the clarifier remained the same and enough plates were added so as to provide a 1 inch spacing between the plates. With this plate addition, sufficient settling area was provided to achieve the original goal of decoupling the clarifier from the thickener.

An operational limitation in the clarifier was encountered while establishing baseline conditions relating to the clarifier design. This limitation will be noted here and discussed in more detail throughout the subsequent discussions. Briefly, the velocity through the clarifier had to be kept below about 0.02 ft/sec. (or 0.12 ft/min.), in order to avoid solids in the clarifier overflow.

The initial goal of the pilot testing at Paducah was to attain a steady-state baseline condition. The underflow concentration attained at Paducah using their standard Dorr-Oliver thickener was usually about 35 wt.% solids. The volume of the pilot unit was about one-third that of the Dorr-Oliver unit. The baseline operational conditions were set so as to yield a 45 wt.% solids concentration in the underflow (i.e., 10 wt.% above the normal underflow concentration). This baseline objective was easily attained and the unit was operated at the same underflow rates, etc. until solid balances were satisfied and no accumulation or depletion of solids within the pilot unit was occurring. Then, the underflow rate was set as low as possible without plugging the underflow line. The solids concentration in the underflow started increasing immediately and increased to as much as 56 wt.%.

The solids concentration in the different streams were measured periodically by both Auburn staff and by TVA staff. The standard procedure developed and used routinely by TVA staff was employed to make these measurements; i.e., measuring the weight loss after two minutes of drying in a microwave oven. This procedure was found to give very good reproducibility and excellent general agreement between measurements made by Auburn staff and those made by TVA staff.

The batch settling behavior of the FGD wastes which were dewatered during the pilot evaluation was determined. A typical batch settling curve for the FGD wastes is shown in Figure 16.2. Initially, batch settling rates were measured for different initial solids concentrations (refer to Appendix G for data), and a standard flux plot was made (Figure 16.3). The settling behavior of the FGD wastes was typical of compressive sludges, similar to those used in the bench scale development work.

Thickener underflow rate and concentration at different times throughout the testing are plotted in Figure 16.4. Initially, between 0 and about 25 hours of operation, the objective was to attain a steady-state baseline. During this time the underflow rate was maintained constant at about 2.8 to 3.0 gpm. The underflow concentration generally increased and leveled off at about 45 wt.% solids.

Throughout the run, the difference between the inlet and outlet mass flow of solids in the dewatering unit was computed. Despite repeated efforts by the TVA staff to calibrate the magnetic flow meters, only rough

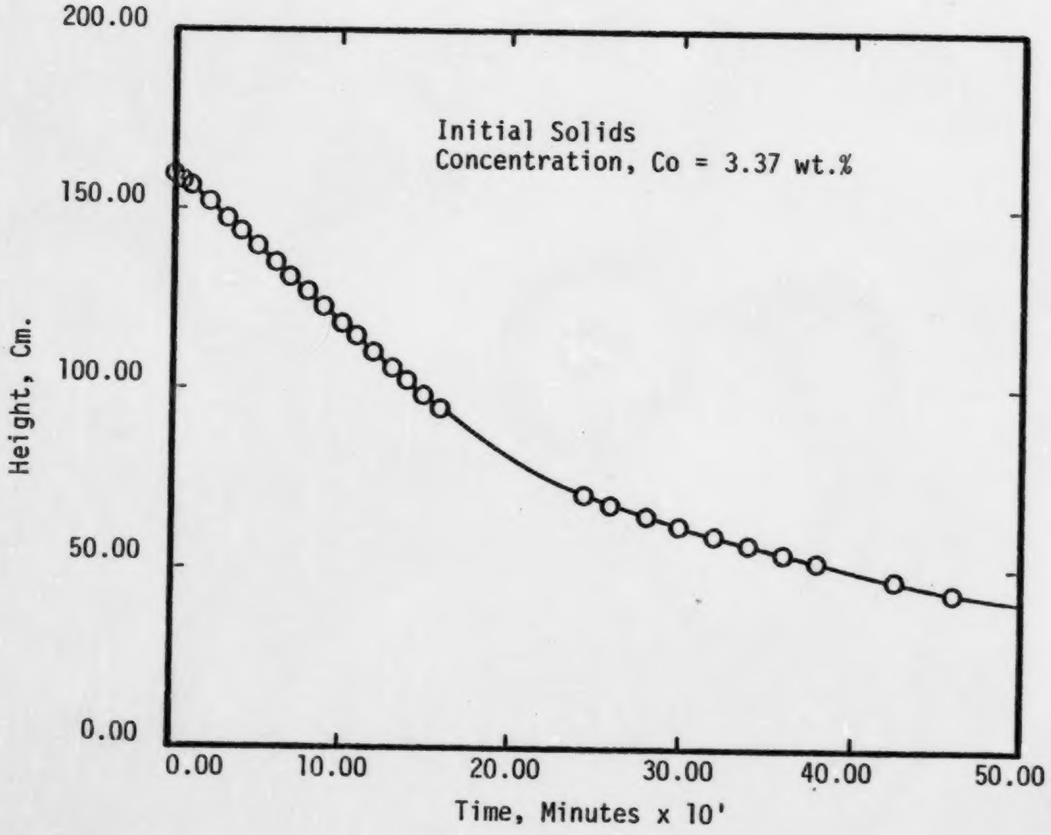


Figure 16.2. Typical Batch Settling Curve for FGD Waste Used In Pilot Evaluation

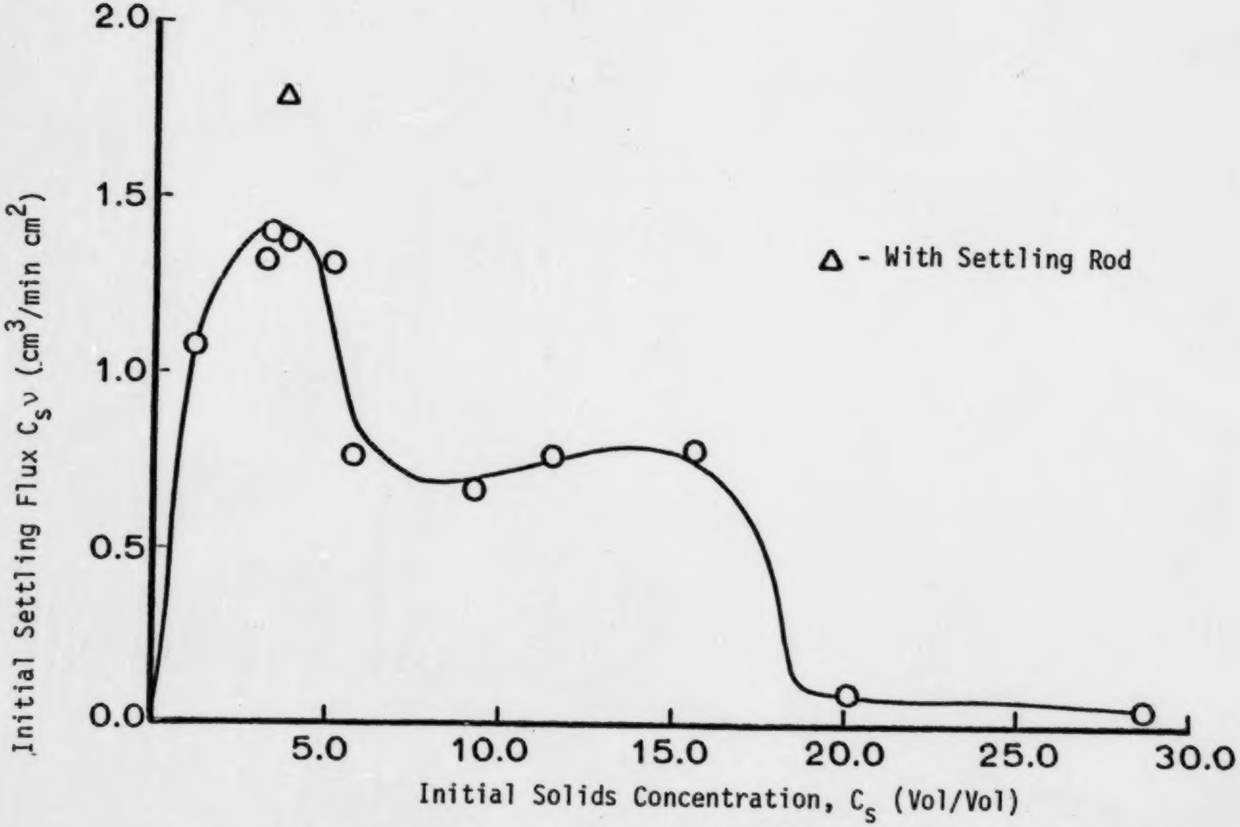


Figure 16.3. Standard Solids Flux Plot Based On Batch Settling Data For FGD Waste (Note: Initial batch settling rates were measured in a 1 liter cylinder; the higher settling rate with the settling rod is typical with hindered settling; refer to laboratory settling studies, Section 5).

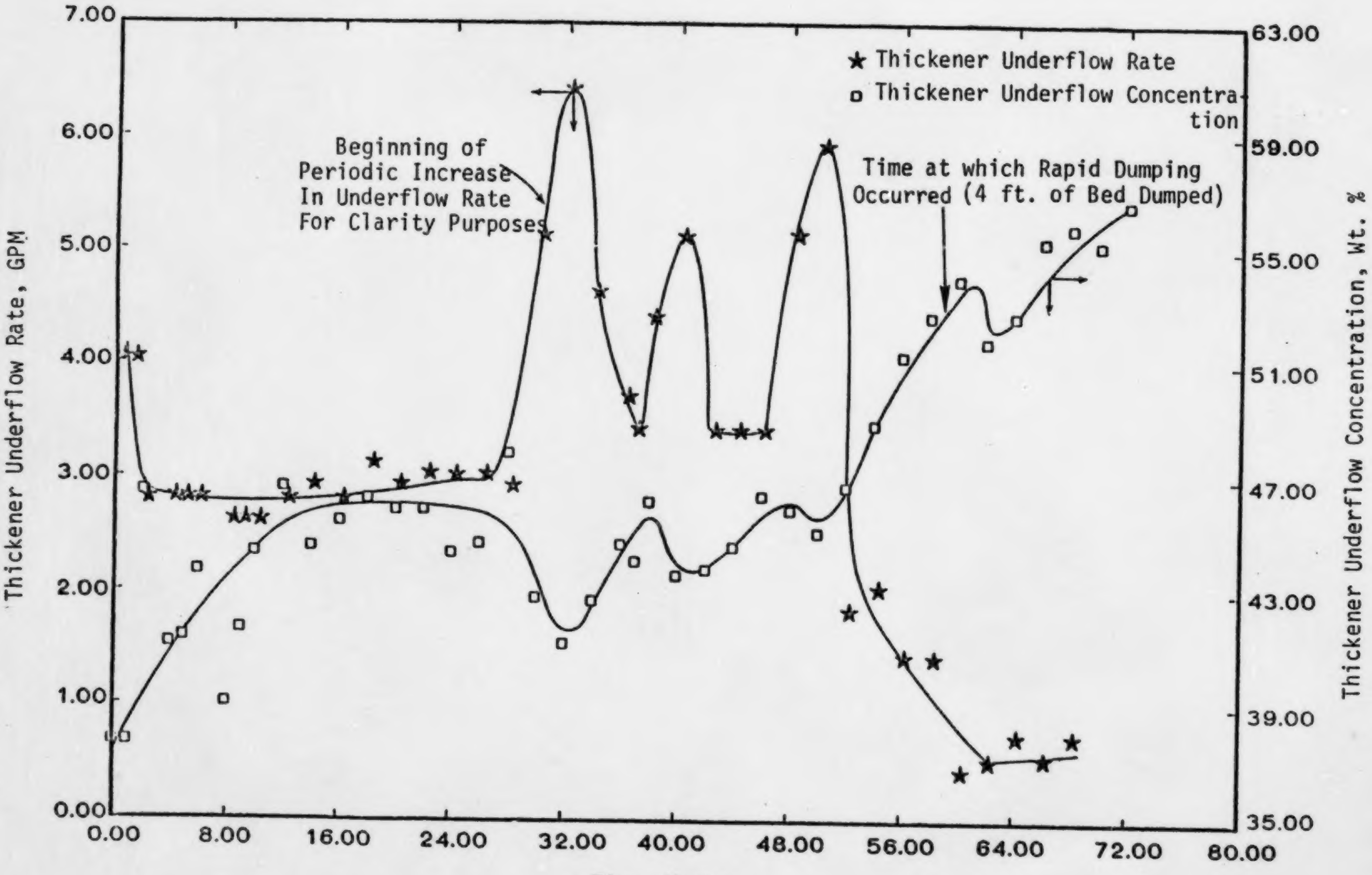


Figure 16.4. Thickener Underflow Rate and Underflow Concentration At Different Times In Pilot Evaluation

balances could be made. As shown in Figure 16.5, solids were being accumulated in the overall system until about 30 hours of operation; between 30 and about 50 hours of operation, the level of solids in the overall system remained about constant. From 50 hours of operation to the end of the test, solids were allowed to accumulate in the system. The objective during this period (>50 hours) was to measure the response in underflow concentration to drastic decreases in the underflow rate.

During the overall steady-state period (30 to 50 hours), the thickener underflow concentration was 44.3 ± 1.6 wt.% solids, and the clarifier underflow concentration was 35.3 ± 1.4 wt.% solids. Thus, there was a relatively small variance in these concentrations (~ 3%) and the overall solids accumulation (7.3 ± 0.2, i.e. 3% variation) during this period. This was true even though the thickener and the clarifier underflow rates varied significantly (4.5 ± 1.1 gpm and 8.9 ± 4.0 gpm, respectively; Figure 16.4 and 16.6).

As outlined in the startup procedure, the thickener was prefilled with sludge prior to starting the run. A major portion of the overall accumulation occurred during the initial part of the test (< 30 hours) and primarily due to filling the clarifier with solids (Figure 16.7). As the clarifier filled with solids, it became necessary to periodically increase the thickener underflow rate in order to not have a turbid overflow. At this time, it was learned that the mass flow of solids in the feed stream to the dewatering unit was being increased periodically to avoid solids accumulation in the scrubber. The periodic raising of the underflow rate from the baseline setting of 2.8 to 3.0 gpm, between 30 and 50 hours of operation, was to offset the periodic increases in solids mass flow in the feedstream. By

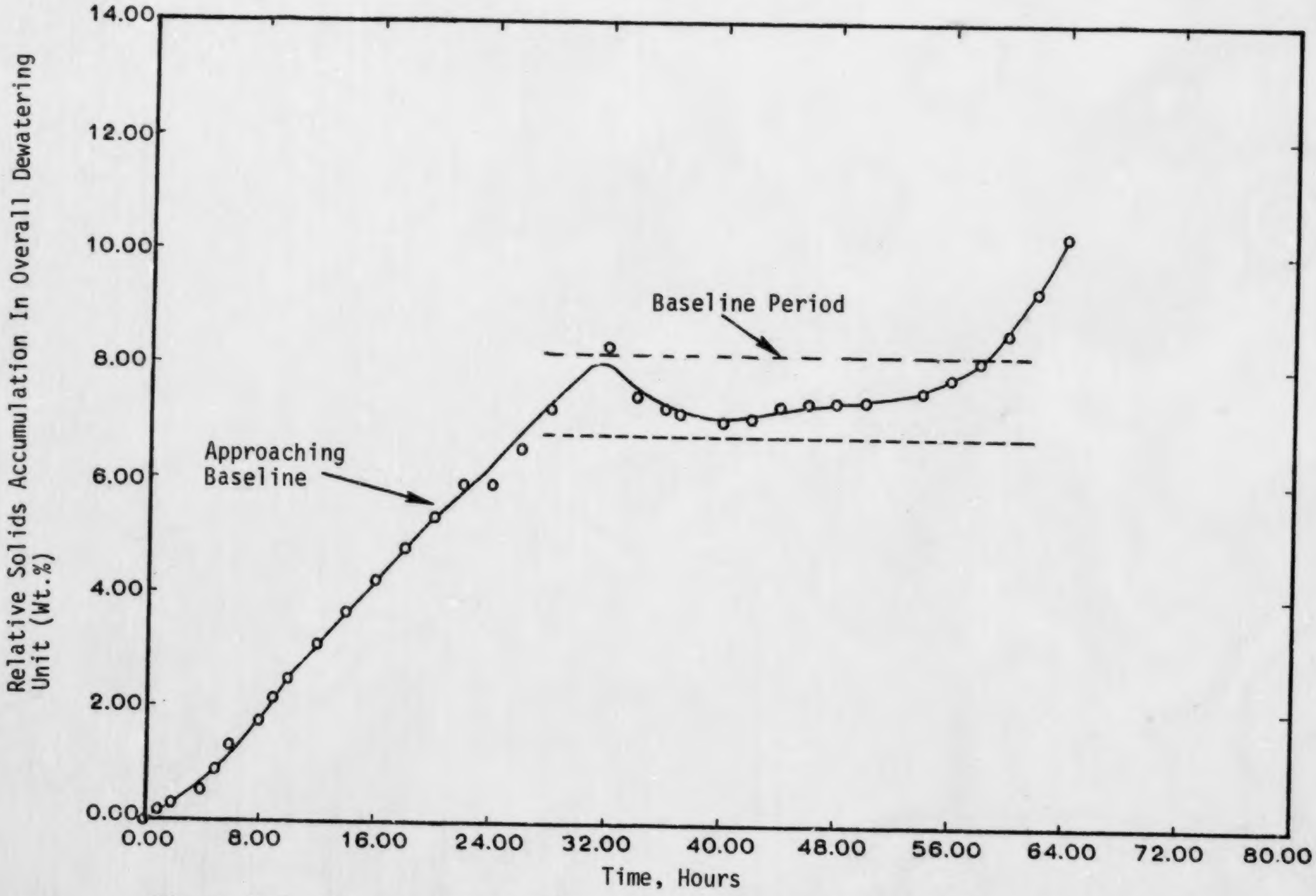


Figure 16.5. Relative Accumulation of Solids In Overall Dewatering Unit At Different Times In Pilot Evaluation

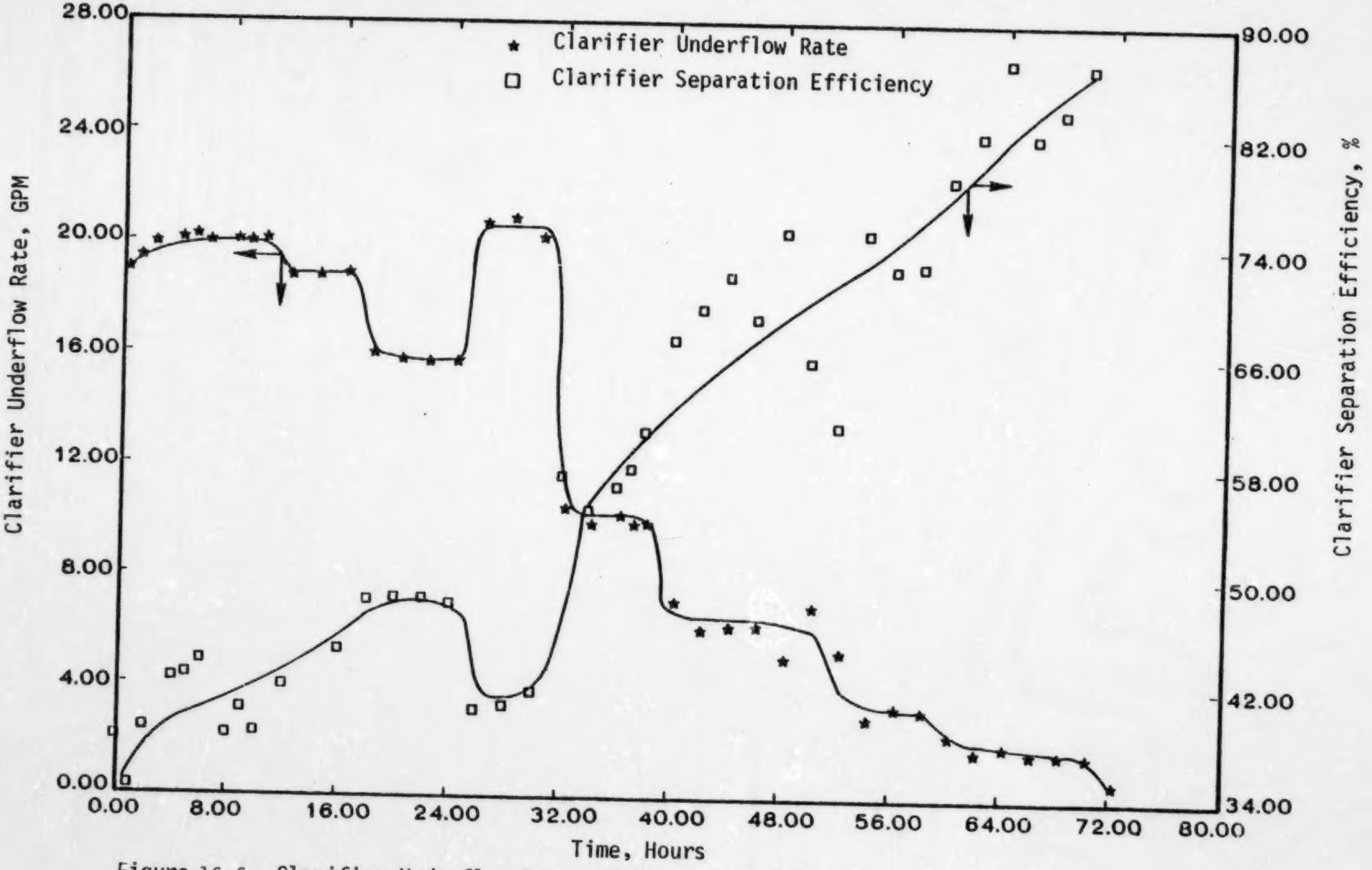


Figure 16.6. Clarifier Underflow Rate and Separation Efficiency At Different Times In Pilot Evaluation

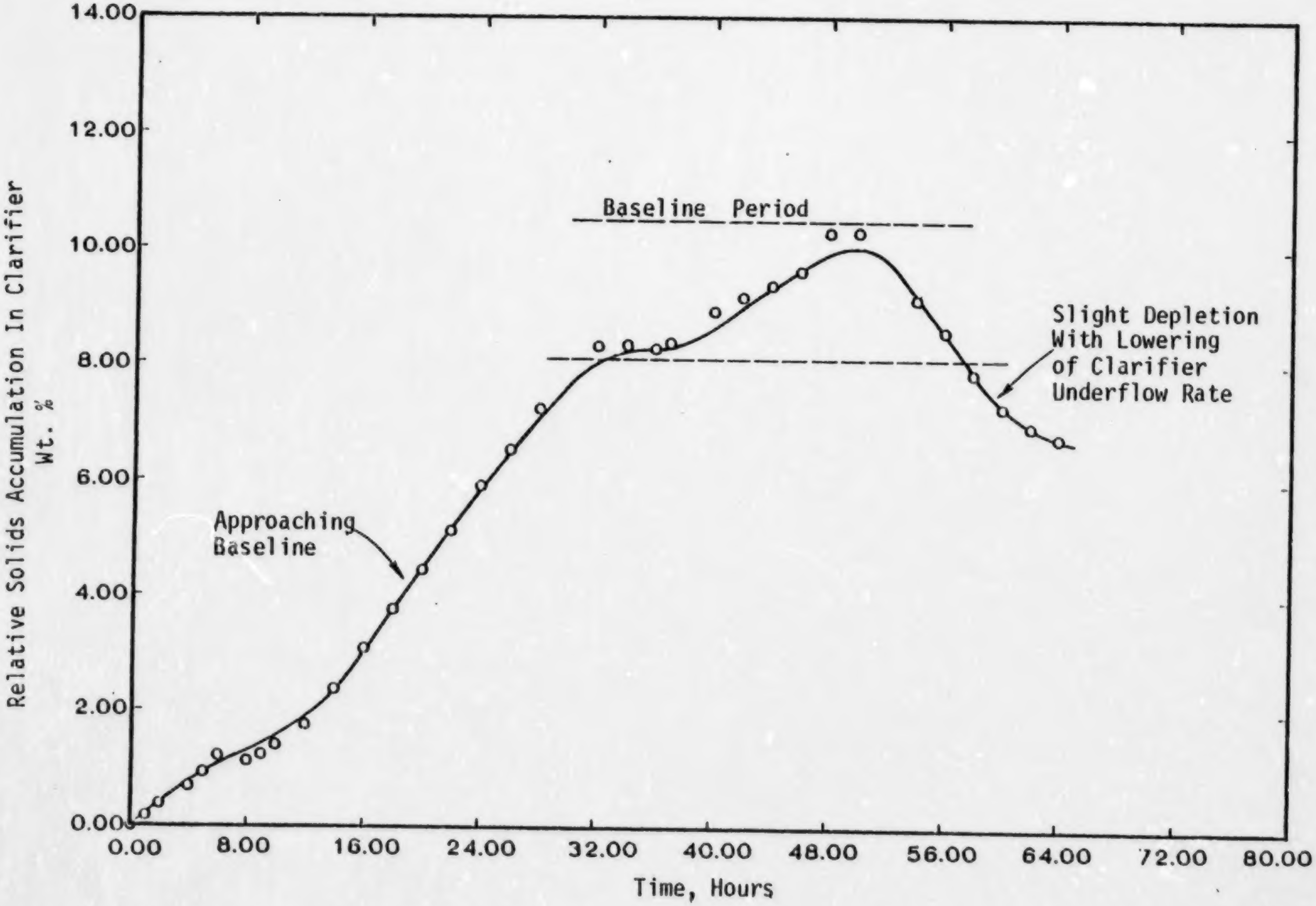


Figure 16.7. Relative Solids Accumulation In The Clarifier At Different Times In The Pilot Evaluation

increasing the thickener underflow rate the net volumetric flow through the upper portion of the clarifier was reduced, and solids elution in the overflow was eliminated. In other words, as pointed out earlier, the velocity through the clarifier had to be kept below a maximum value (~ 0.02 ft/sec. or 0.12 ft/min.).

Once this operational limitation was discovered, a splitter was used to regulate the volumetric flow of the feedstream so that the underflow rate would not have to be adjusted periodically for overflow clarity.

During a period of 30 to 50 hours, each time the underflow rate was increased, the underflow concentration decreased to some extent. After the feedstream splitter was added (> 50 hours), the underflow rate was decreased drastically. In response, the underflow concentration increased drastically. The underflow concentration increased to as high as 56 wt.%.

At high concentrations, the underflow was very viscous, but it was pumpable even when its concentration was 56 wt.%. Like many solid concentrates the thickener underflow was thixotropic, and to prevent it from setting-up, it was always subjected to some form of movement or shear. The time allocated for the pilot testing did not permit operation at underflow concentrations higher than 56 wt.%. Operation above this concentration would have been difficult anyway because of the underflow's high viscosity and tendency to plug.

Unlike the thickener, when the clarifier underflow rate was drastically reduced, its underflow concentration increased only slightly on the average (Figure 16.8). This is to be expected since little compression settling occurs in the clarifier.

The function of the clarifier was to produce as much clear overflow as possible, whereas that of the thickener was to produce as concentrated

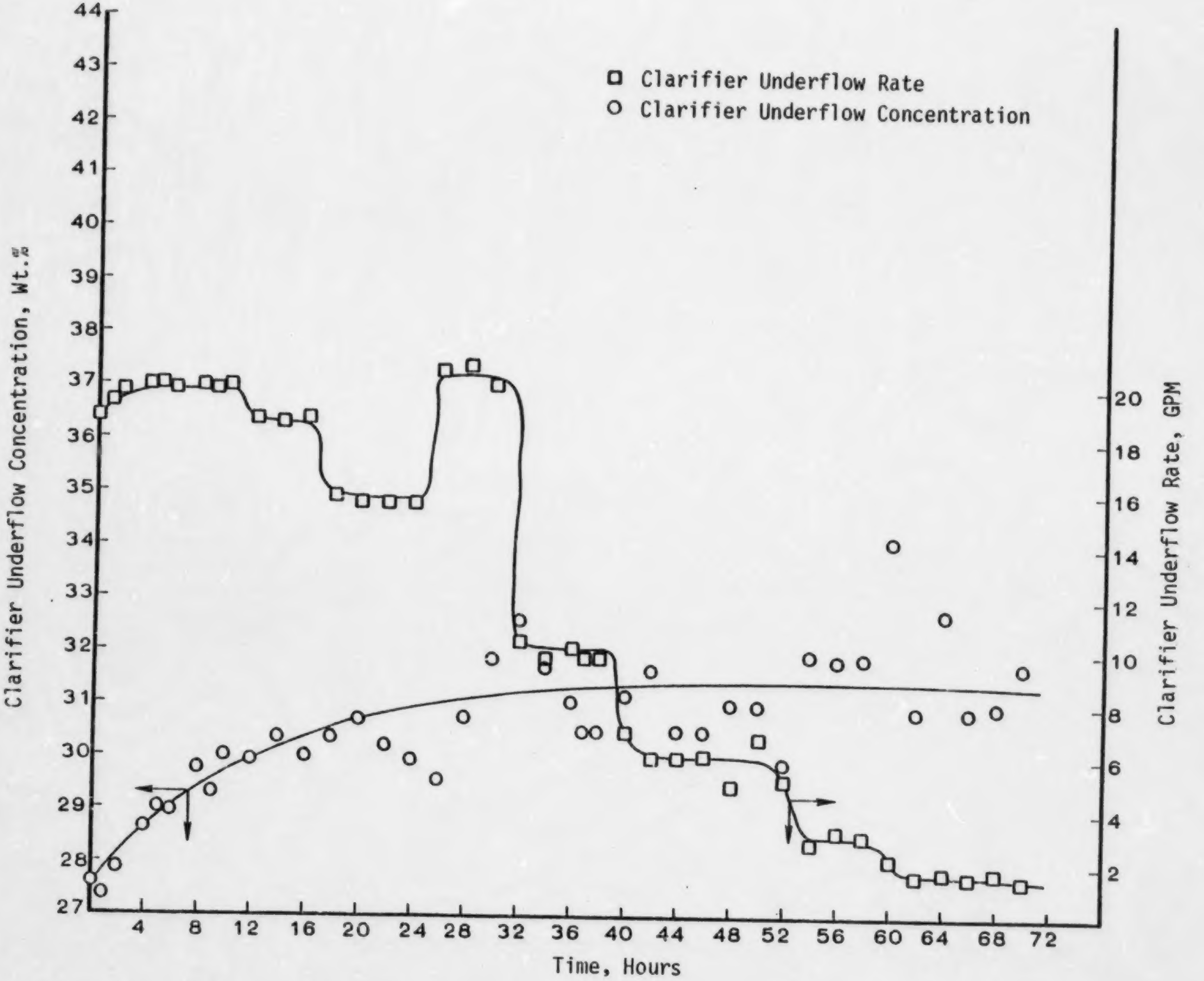


Figure 16.8. Clarifier Underflow Concentration and Flowrate At Different Times During Pilot Evaluation

an underflow as possible. Consequently, a reasonable measure of the clarifier's performance is the efficiency by which it dewatered the feedstream. The clarifier's separation efficiency was defined as the percentage of the water entering in the feedstream that is separated from it and leaves in the clear overflow. A plot of separation efficiency versus underflow rate is shown in Figure 16.6. After the initial 8 to 10 hours of operation, each time the underflow rate was decreased there was a corresponding increase in separation efficiency.

As the clarifier underflow rate was decreased, the solids concentration in the thickener decreased. Decreasing the clarifier underflow rate results in a lowering of the upward velocity in the upper portion of the thickener. Apparently there was less mixing of the concentrated clarifier underflow and the feedstream; thus, a less concentrated thickener overflow. At the very low clarifier underflow rates (2.0 gpm) some clarification occurred in the thickener and its overflow concentration was less than that of the feedstream (Figure 16.8).

As the clarifier underflow rate is decreased, its tendency to plug increases. It thus appears that the clarifier underflow rate should be set as low as possible without having to be concerned about is plugging.

During the latter period of the run, when the thickener underflow rate was drastically reduced from its baseline setting (2.8 to 3.0 gpm), a mistake was made which allowed a most interesting observation. After a sample of the underflow was taken (at 58 hours of operation), the sampling port was left fully open for several hours. Once discovered, the sampling port was immediately closed. It was estimated that about 4 feet of concentrated solids from the bottom of the thickener were dumped while the sampling port was left open. A sample of the underflow was collected immediately after this dumping; it was determined to be within one percent of the underflow concentration before the dumping occurred (i.e., 52.7 wt.%). If the underflow rate had been kept at the high rate at which the dumping occurred, the underflow concentration would have eventually dropped to a low value. Apparently, by reducing the underflow rate back to its original value before completely dumping the concentrated solids, the underflow concentration decreased only slightly.

It is well known that the solids concentration profile in the thickening zone for compression settling is very flat (ref. 21; Appendix A). Often, this is interpreted to mean that the residence time of solids in the thickening zone is not important, and this in turn is misconstrued to imply that the underflow rate has little effect on underflow concentration. It is well known, however, that the addition of a settling basin to the bottom of a laminar-type clarifier, to allow compression settling to occur, results in a more concentrated underflow. A laminar-type clarifier consists of settling plates stacked on top of one another, as is true for the one used here. As a result, it does not provide any significant sludge blanket depth to allow compression settling to occur. The addition of a settling chamber to the bottom of such a clarifier allows the settling sludge to compress itself. Also, as experienced in this test run, decreasing the thickener underflow rate, results in a more concentrated underflow (Figure 16.4). In summary, it appears that a zone of almost constant concentration accumulates in the bottom of the thickener, and the concentration in this zone varies inversely with the underflow rate.

Section 17

INTRODUCTION OF THE CONCEPT OF BANG-BANG OPERATION

Assuming the observations made during the final pilot plant tests are valid, then the optimal operation of a solid-liquid separator would appear to be "bang-bang" operation. Bang-bang operation, or control, is standard terminology used to designate cyclic "on-off" operation of a flow stream. More generally speaking, during operation of any processing system, a stream is set to flow at one rate for a certain portion of a cycle and then the flow-rate is set at a different magnitude for the remainder of the cycle (Figure 17.1). Typically, the flow rate is negligible for a certain portion of the cycle (i.e., the valve is mostly closed), and then it is stepped up to maximum flow for the remainder of the cycle (i.e., the valve is completely open). The term bang-bang ascribes to the banging of the valve fully closed and fully open.

Since the underflow concentration, Cu, varies inversely with respect to the underflow rate, U, to maximize Cu, U should be set as low as possible. (Note: U and Qu are used interchangeably for underflow rate in this report). U cannot usually be zero because most difficult-to-separate sludges like FGD sludges are highly thixotropic and their viscosity usually increases drastically without any applied shear (i.e., no pumping). To prevent the underflow sludge from setting-up and plugging (e.g. forming cement) a certain minimum flow rate (U_m) or shear rate is generally required.

In summary, the bang-bang operation should be carried out as follows: The dewatering system should be operated at U_m, the minimum

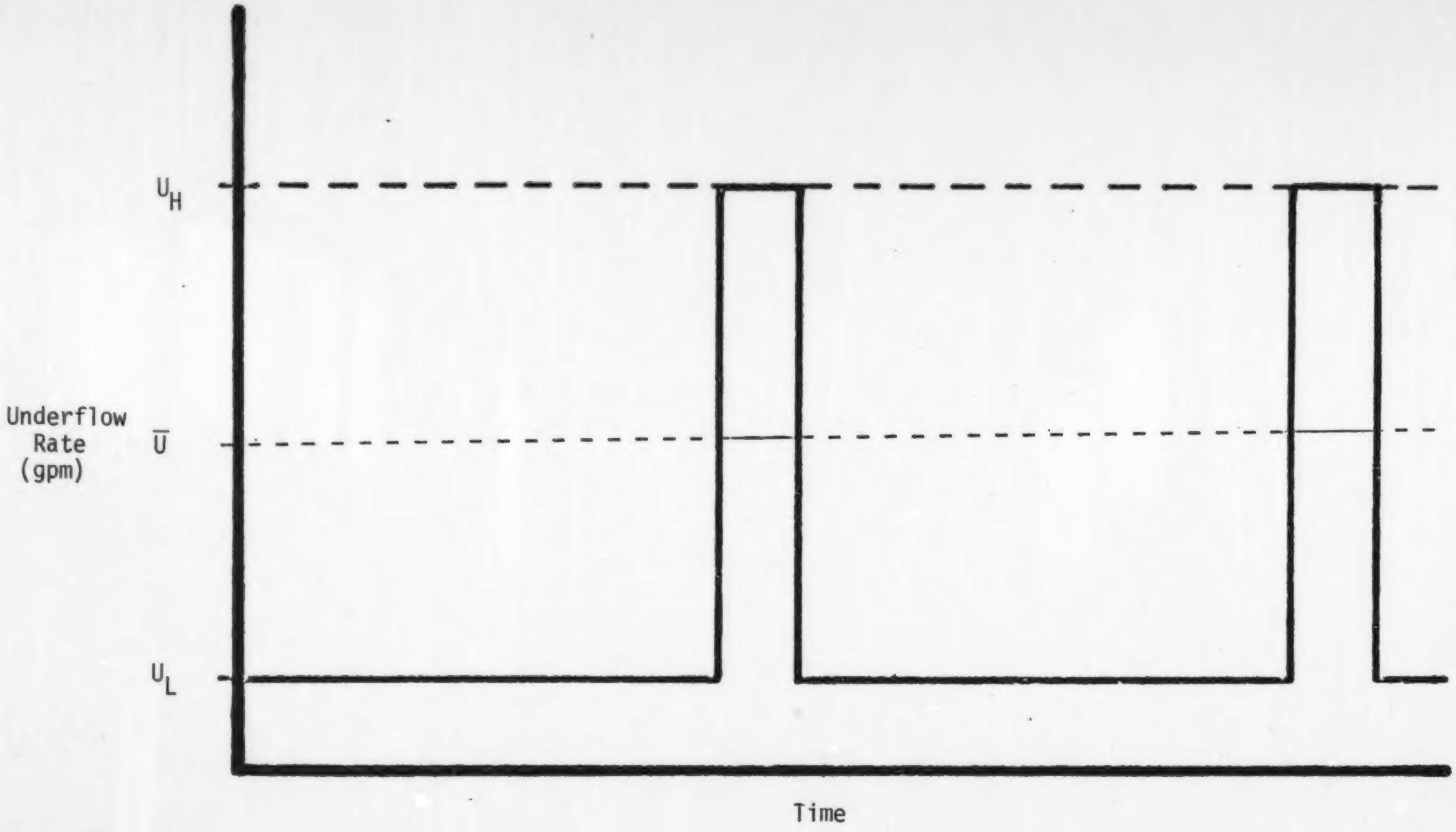


Figure 17.1. Underflow Rate Versus Time for a Typical "Bang-Bang" Operation

underflow rate required to prevent plugging. In order to maintain a balance between solids entering and leaving the system, the deep-bed thickener should be periodically dumped by shifting to a high underflow rate U_H as illustrated in Figure 17.2. This is done such that, during dumping ($U = U_H$), the sludge blanket would never be depleted to the extent that a portion of the essentially constant concentration zone (illustrated in Figure 17.3) would disappear completely. Even during the dumping portion of the cycle, shown in Figures 17.1 and 17.2, the underflow concentration would remain essentially equal to C_m.

If U is set equal to U_m, then Cu should be at its maximum value and the overall separation efficiency would be maximized. For most thickening systems, U cannot be set equal to U_m because maintaining the clarity of the overflow would become a problem. The sludge blanket would gradually increase in depth, and the overflow from the thickener would soon become cloudy (i.e., the thickener would flood with solids). For the dewatering system tested here, this limitation was overcome, allowing U to be set at U_m.

If the settling area (i.e., the cross-sectional area of the thickener) is sufficiently large, the thickener will not flood when the underflow rate is set low. So, why not make the cross-sectional area of the thickener very large and set the underflow rate equal to U_m? Actually, this is what is often done when sufficient land area is available so that the sludge can be ponded. Settling ponds, however, are not desirable in numerous cases for a variety of reasons, as is assumed to be the case here.

The question naturally arises as to why not operate at some mean flow rate U instead of resorting to the complicated bang-bang operation

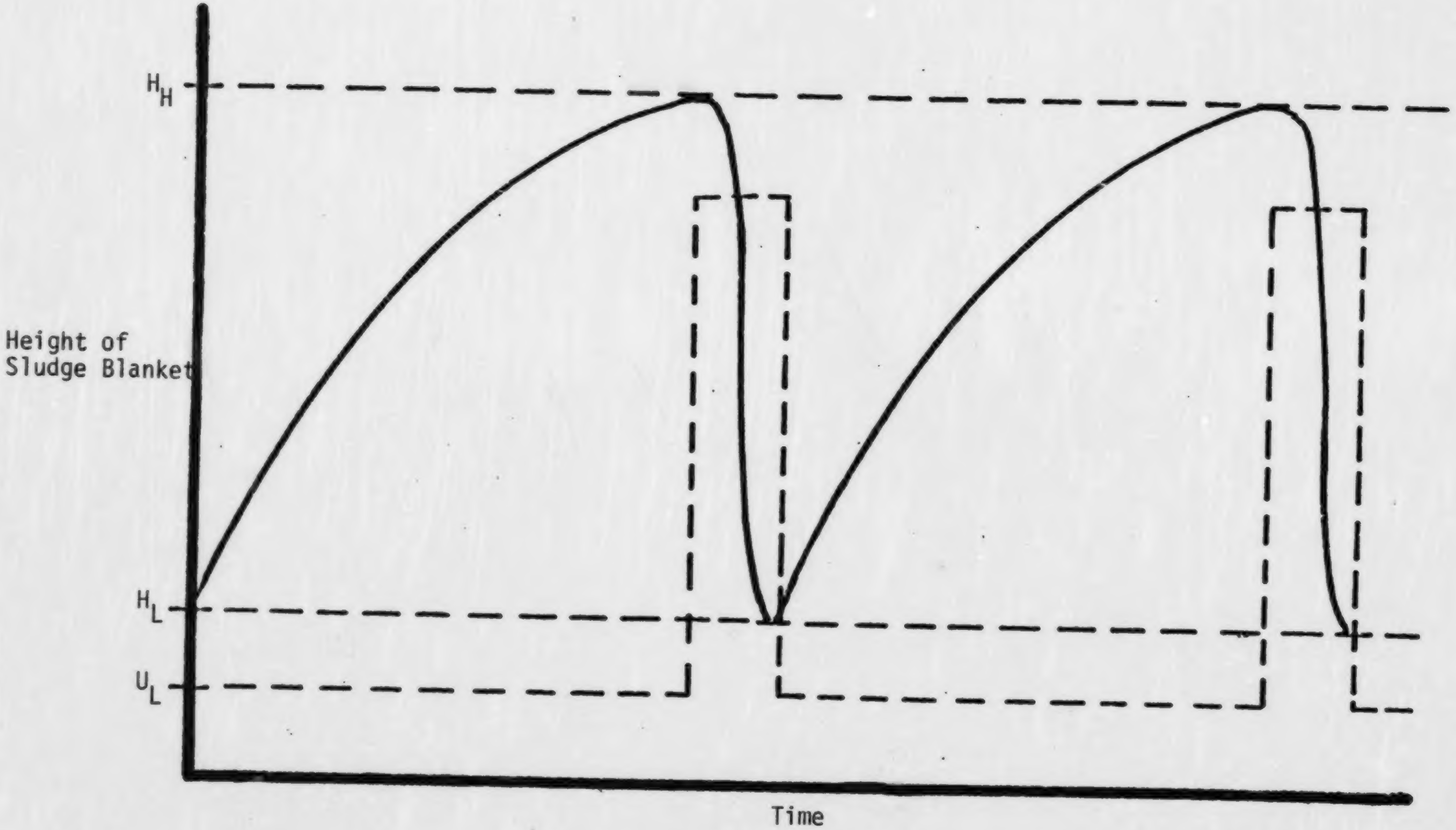
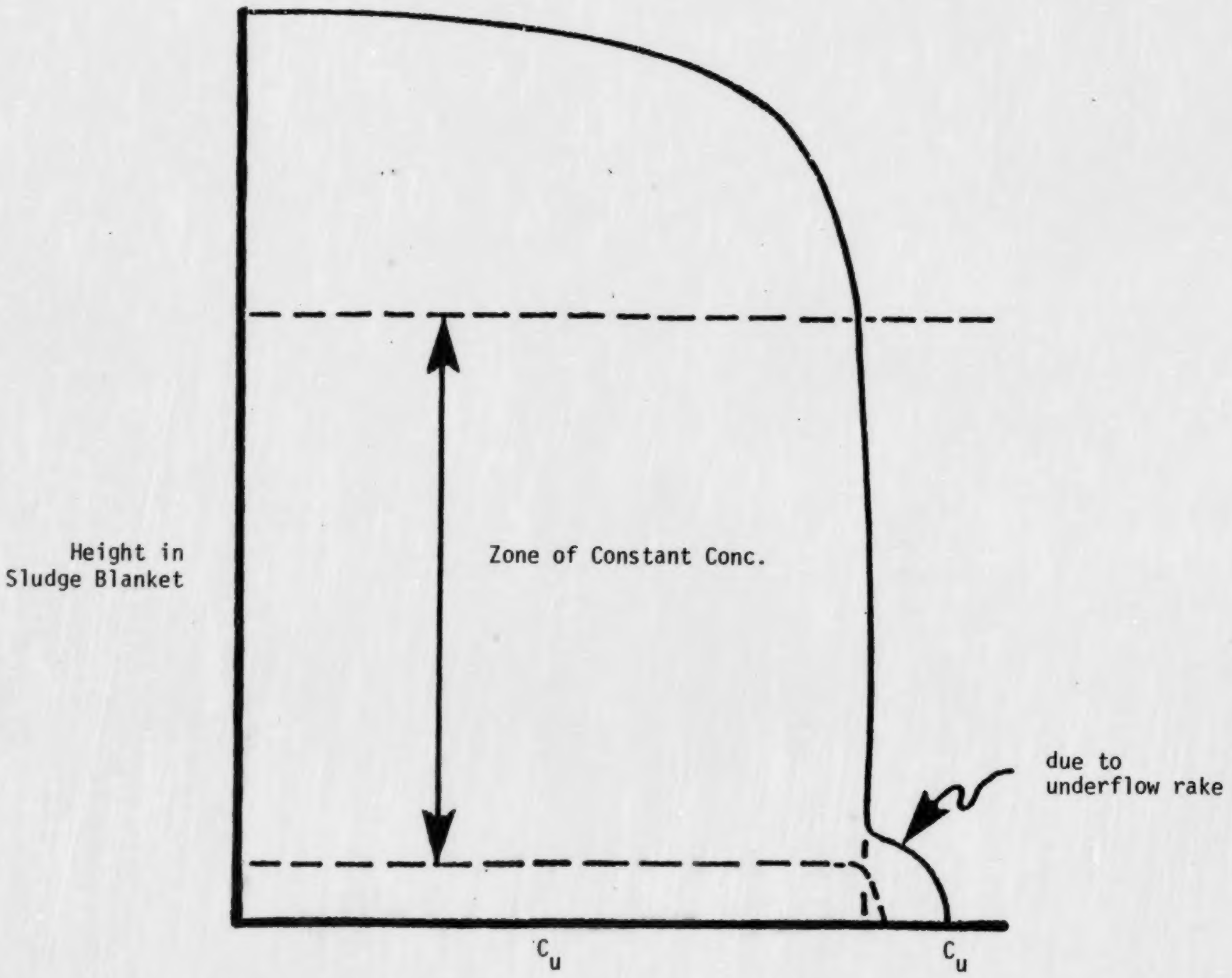


Figure 17.2. Response of Sludge Blanket Height to Change in Underflow Rate with Respect to Time.



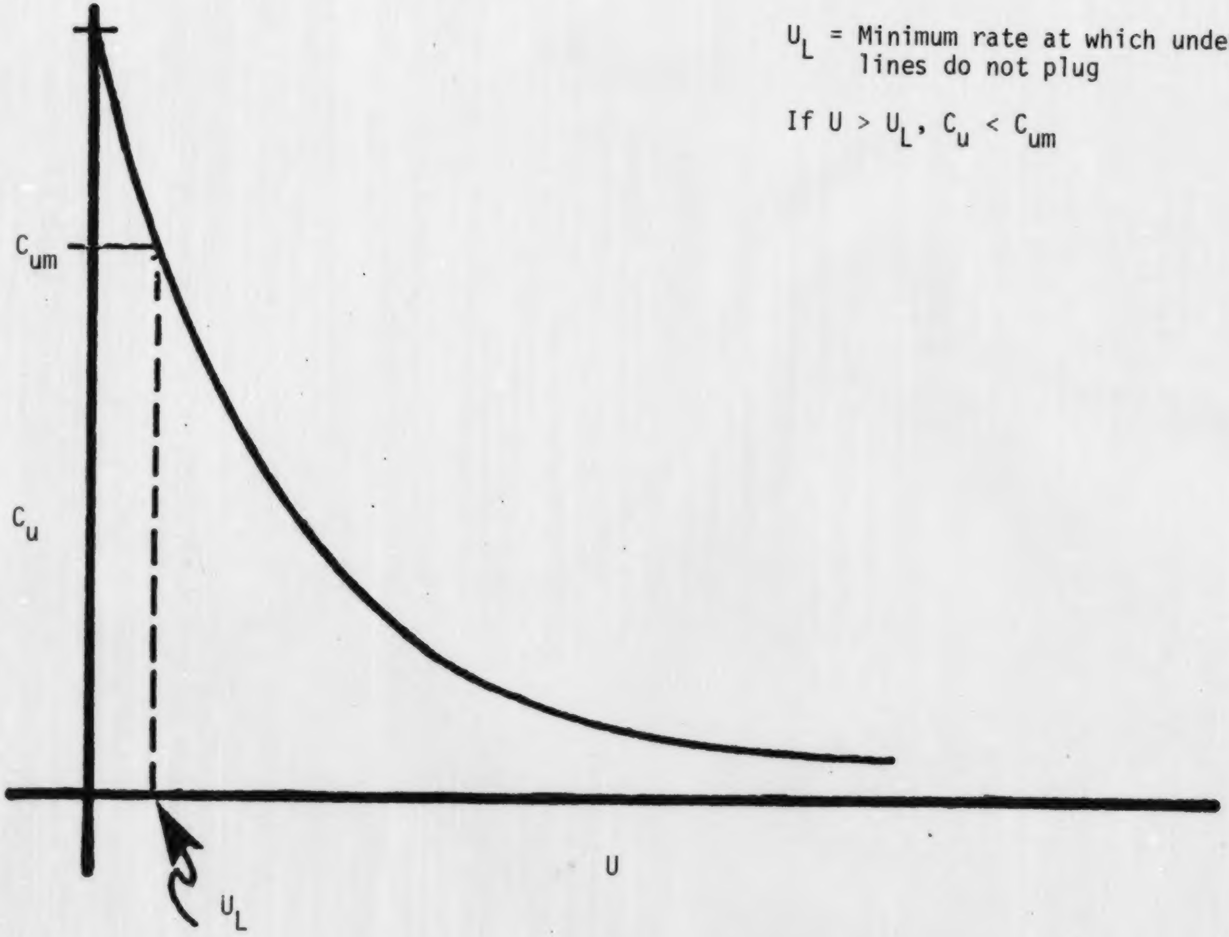
Height in
Sludge Blanket
Zone of Constant Conc.
due to
underflow rake
C_u
C_u

C_u = Underflow Conc. = f(underflow rate, U_L)

Figure 17.3. Sludge Blanket Height Versus Concentration in Sludge Bed, C

described above. Apparently, by setting U equal to U_m for an extended portion of the operating cycle, sufficient solids residence time is allotted so that the layers of settling solids in the bottom portion of the sludge blanket can be subjected to sufficient compressive forces to attain a maximum underflow concentration C_{um}. That is, the compressive force exceeds some resistive yield strength (τ_{om}) associated with attaining C_{um}. Layers of sludge having a concentration of C_{um} appear to propagate rapidly upward in the sludge blanket yielding a zone of essentially constant concentration (C_{um}) as illustrated in Figure 17.3. (The exact mechanism involved here is not well understood; to obtain more insight into this mechanism, further research is required). In summary, by setting $U = U_m$, C_u increases to C_{um} and maximum is achieved; by rapidly dumping at $U = U_H$, the solids mass balances around the system are satisfied with little or no sacrifice in separation efficiency (C_u remains equal to C_{um}). Conversely, if U is set at some mean value $\bar{U}$, then the underflow concentration remains at some value lower than C_{um} and maximum separation is not achieved (Figure 17.4).

The concept of bang-bang thickener operation described here is completely new; further research is required to substantiate the concept.



U_L = Minimum rate at which underflow lines do not plug

If $U > U_L$, $C_u < C_{um}$

Figure 17.4. Concentration in Zone of Constant Concentration vs Underflow Rate

Section 18

OVERALL CONCLUSIONS AND RECOMMENDATIONS

- It is possible using the dewatering system proposed herein, to, in effect, decouple the clarification function of an FGD waste dewatering system from its thickening function.

- Using the proposed dewatering system, it is possible to concentrate unoxidized FGD wastes to a high solids concentration (~56 wt.%) without the addition of additives and with the use of reasonable equipment sizes.

- By manipulating the clarifier underflow rate, plugging in the clarifier can be avoided and high underflow concentrations can still be attained by operating at a relatively low thickener underflow rate, (note: This feature of the proposed dewatering system is especially advantageous when applied to the dewatering of oxidized FGD wastes. Overdesign of the clarifier which often results in plugging problems in the underflow lines, would not be a problem and underdesign of the clarifier could be offset with the addition of the thickener).

- When operating the dewatering system, the thickener should be operated in a "bang-bang" mode; i.e., the underflow rate should be set (for most of the "bang-bang" cycle) as low as possible without plugging the underflow lines. The clarifier underflow rate should also be set as low as possible without having to be concerned about plugging. Upon attainment of a maximum possible underflow concentration the underflow rate should be stepped up, for a brief period, in order to remove the underflow.

- In designing the clarifier, its cross-sectional area should be sufficient so that the velocity through its settling section is less than about 0.02 ft/sec.

Section 19

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Appendix A

"A Model for Continuous Thickening"

Fundamental Aspects of Hydrometallurgical Processes, AIChE Symp. Series, 173, Vol. 74, 1978, pp. 67-74. (Published with permission.)

A MODEL FOR CONTINUOUS THICKENING

A. R. Tarrer
D. L. Vives
and
D. M. Kennedy

The effect of two major operational parameters, solids residence time and depth of the sludge blanket, on thickening capacity are often ignored during design and analysis of a thickener, mainly because of lack of satisfactory means to do so. A well-known theory for describing batch sedimentation (Kynch's theory) is utilized here as a basis for characterizing different sludge systems within dilute concentration ranges, where appreciable compressive settling does not occur. Guided by observation (*1*), the settling process occurring in a continuous thickener is modeled by linearly superimposing the batch settling mechanism, as described by Kynch's analysis, onto a plug-flow fluid pattern. The result is a simple method by which the effects of the depth of the sludge blanket and the residence time of solids can be estimated for systems where compression does not occur to an appreciable extent.

CHARACTERIZATION PROCEDURE

THEORY

Kynch's analysis for batch settling is based on the simplifying assumptions that rate of descent of the interface layer v is a function solely of local solids concentration $v = v(C_s)$ and that a continuous medium exists throughout the subsiding sludge blanket. Once the solids become sufficiently close that channeling or extensive bridging together of solids occurs (Figures 1*a* and *b*), these assumptions, and thus Kynch's analysis, are obviously not valid. Also, once compression, a state in which the solids in adjacent layers are essentially lying on top of one another, is reached, v is not $v(C_s)$, but depends on other parameters, particularly height [that is, $v(H, C_s)$]. By limiting application to lower concentration ranges where the above assumptions are valid, Kynch's analysis can be used to describe batch settling, thereby offering a means for characterizing different sludge systems according to parameter values in the settling rate expression $v = v(C_s)$.

Standard practice is to use batch settling tests in characterizing the settling behavior of a sludge system. During a batch settling test, a column of sludge initially at uniform concentration is allowed to settle. Normally, throughout a settling test, a clear line of demarcation exists between settling solids and supernatant. The subsidence of this sludge-supernatant interface is measured as a function of time (Figure 3). Typically, the resulting curve has an initial linear section, corresponding to the subsidence of solids at the initial test concentration. Once the concentration gradient propagated from the bottom of the batch column reaches the interface, the settling curve discontinuously bends, exponentially approaching some final equilibrium height. A series of batch settling curves covering a range of initial test concentrations are usually the only data available for characterization, as is assumed to be the case here. Modeling of batch settling to facilitate characterization thus entails predicting the height of the interface at different settling times or, as shown later, when successive concentration layers reach the interface.

According to developments by Talmadge and Fitch (*11*) based on Kynch's analysis, a concentration layer propagates upward from the bottom of the settling column at the following rate:

$$w = -d(C_s v)/dC_s \qquad (1)$$

Thus, the height of the interface H at any specified time is a function only of C_s since

$$H = H_o - vt, \; t < t_k$$

and

$$H = wt = -[d(C_s v)/dC_s]\, t, \; t \geqslant t_k \qquad (2)$$

where t_{kl} is the time at which the concentration gradient reaches the interface and H_o is the initial interface height.

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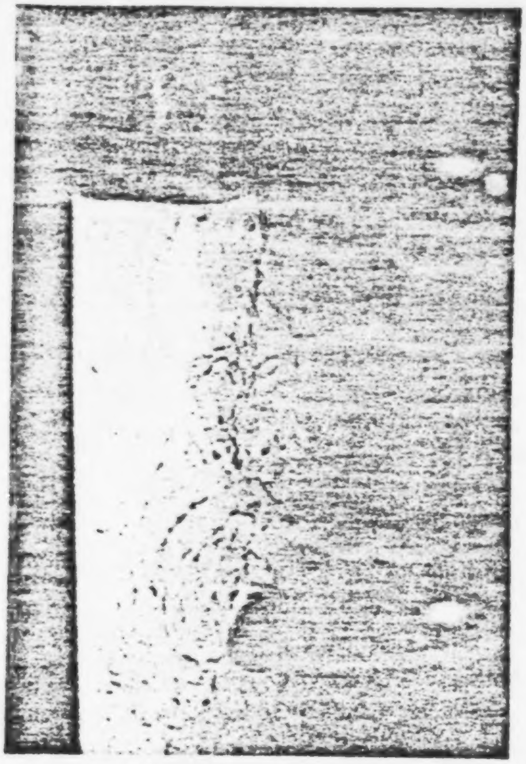

Fig. 1*a*. Large channel formed around sample port carrying water out of sludge blanket.

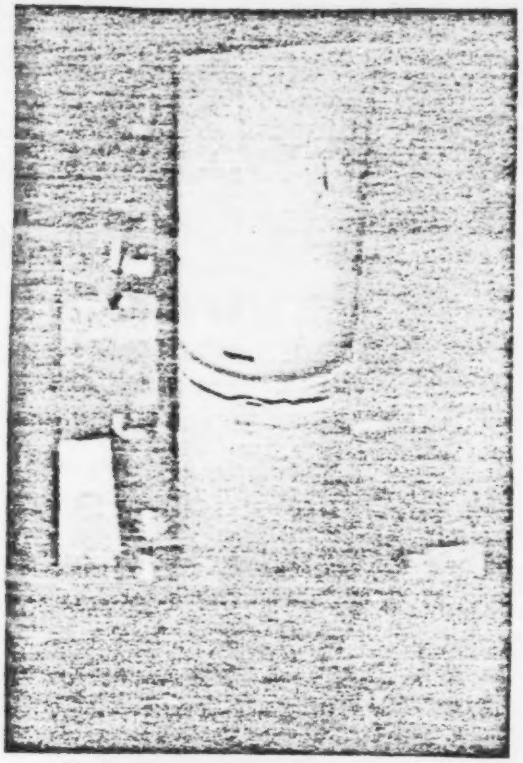

Fig. 1*b*. Close-up of discontinuity in settling test.

The time t_{ki} at which the constant velocity period ends is given by

$$H_o - v\big|_{C_{so}} t = -t_k \left.\frac{d(C_s v)}{dC_s}\right|_{C_{sk}}$$

or

$$t_k = -H_o/[C_{so}(dv/dC_s)\big|_{C_{sk}}] \qquad (3)$$

C_{so} is the initial solids concentration in a column with cross-sectional area A filled to a height H_o. C_{sk} is the concentration of the interface layer existing at the time that the discontinuity occurs in the settling curve, that is, at t_{ki}, and, according to Kynch, can be calculated using the following relationship:

$$(C_{so}v_o - C_{sk}v_k)/(C_{sk} - C_{so}) = -d(C_s v)/dC_s\big|_{C_{sk}} \qquad (4)$$

When a concentration layer reaches the interface, all solids must have settled through it since it was propagated from the bottom. That is, for $t \geqslant t_k$

$$C_{so}H_oA = C_sAt(v + w)$$

or

$$-C_s^2 t(dv/dC_s) = C_{so}H_o \qquad (5)$$

Thus, it is seen that given $v = v(C_s)$, C_{sk} and t_{ki} may be calculated using Equations (4) and (3), respectively; also, for $t > t_{ki}$, C_s may be computed using Equation (5); and subsequently, H, using Equation (2). In other words, the entire batch settling curve becomes determined, given $v = v(C_s)$ and initial conditions (C_{so}, H_o).

After Figure 2 was analyzed as presented by Dorr-Oliver (*4*), the following empirical form was chosen for $v = v(C_s)$:

$$v = B_o \exp(B_1 C_s + B_2 C_s^2) + B_3 \qquad (6)$$

The settling behavior of a particular sludge is thus characterized by the set of empirical parameter values B_i, $i = 0, 1, 2, 3$.

The technique employed for the selection of these values was a curve fit wherein the batch settling equations [Equations (2) to (5)] are fitted simultaneously to a series of batch test data so as to minimize the sum of squares deviation between actual and computed interface heights. It should be noted that B_3 is fixed to ensure that v will be zero at C_{sf}, which is defined by

$$C_{sf} = C_{so}H_o/H_{inf} \qquad (7)$$

EVALUATION

Glass beads settle almost as discrete particles with little interparticle bridging occurring. For this reason, glass bead data collected by Stroupe (*10*) were chosen for initial testing of the proposed procedure for determining the four settling param-

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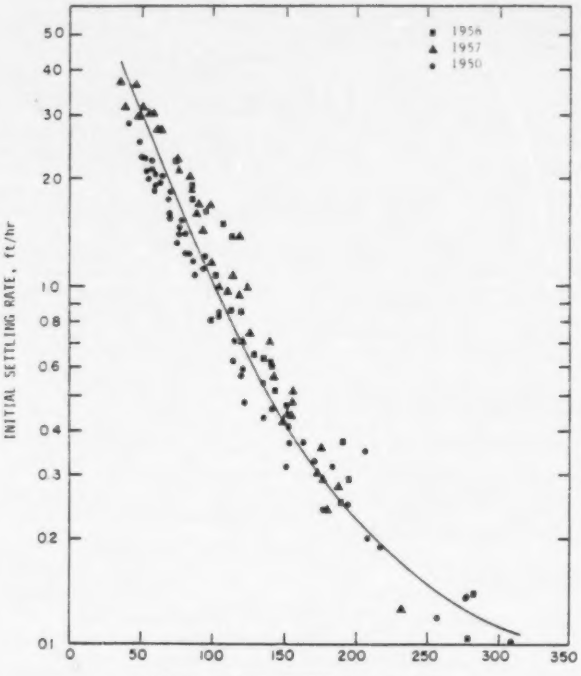

Fig. 2. Settling rate as a function of concentration (*4*).

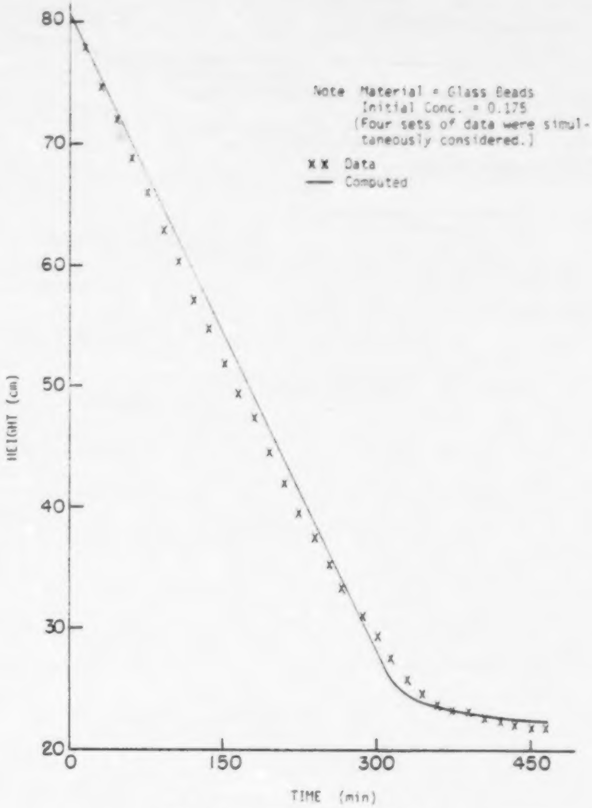

Fig. 3. Interface height vs. time.

eters. Four sets of batch test data were fitted simultaneously; in so doing, the search routine (*8*) converged in a reasonable amount of computer time (less than 1 min on a CDC 6500). Figure 3 shows one plot of interface height vs. time data obtained by Stroupe (*10*) compared with the fitted curve. This fit is representative of those obtained with the other three initial conditions and was taken as sufficient verification of the algorithm used to implement the proposed procedure.

To evaluate Kynch's analysis for a less ideal system (one in which significant interparticle bridging does occur), using calcium carbonate as a test material, concentration profiles in a batch settling column (Figures 5 and 6) were measured; the proposed characterization procedure was applied to determine values for the settling parameters; and theoretical and measured concentration profiles were compared.

In searching for the settling parameters, the search routine (*8*) would not converge unless guess and final values were close; randomly chosen guess values could not be used. As a result, a computer routine was developed to locate, as reasonable estimates, values for B_i, $i = 0, 1, 2$ [B_3 being determined by Equation (7)] so as to fit Equations (2) to (5) exactly for three data points in the curved region of a settling curve. From these estimates as initial (or guess) values, rapid convergence was attained. Figure 4 compares the interface height vs. time data with the fitted curves. In determining the associated settling parameters, five sets of batch settling data were fitted simultaneously, and an even number of uniformly distributed (at 3 h intervals) data points in the linear and curved sections

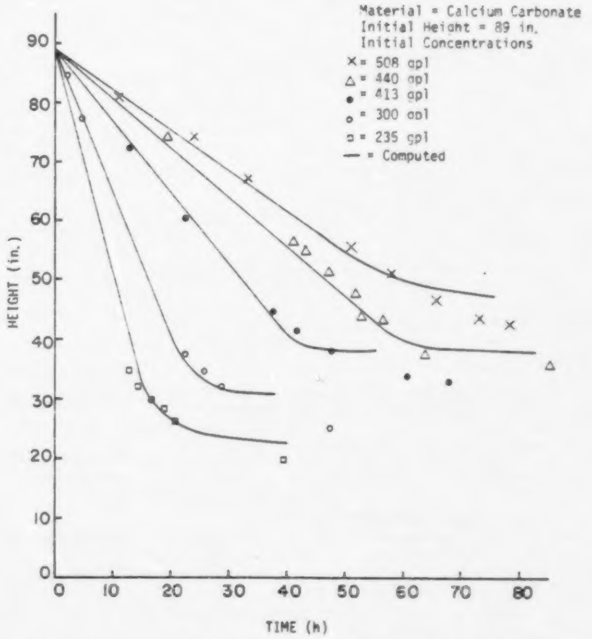

Fig. 4. Batch interface subsidence curves for calcium carbonate.

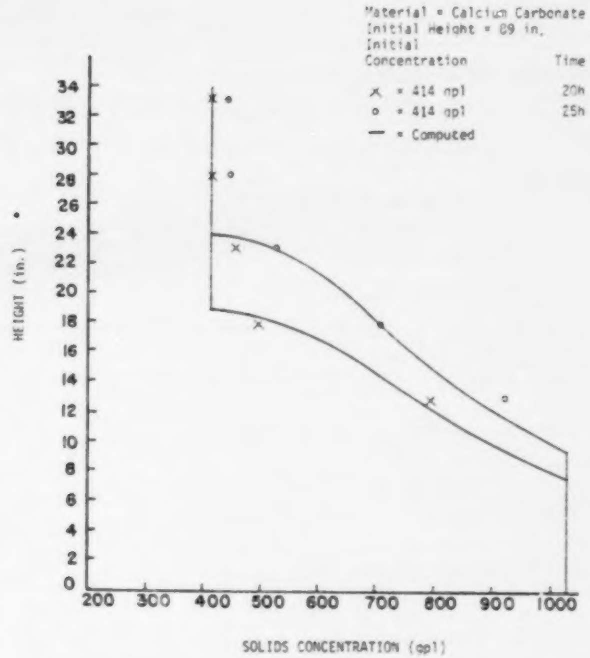

Fig. 5. Batch concentration profiles for calcium carbonate.

of the settling curves were used. However, the data points in the curved section were weighted heavily (a weighting factor of 100 was used); by so doing, better agreement resulted between predicted and measured concentration profiles (Figures 5 and 6).

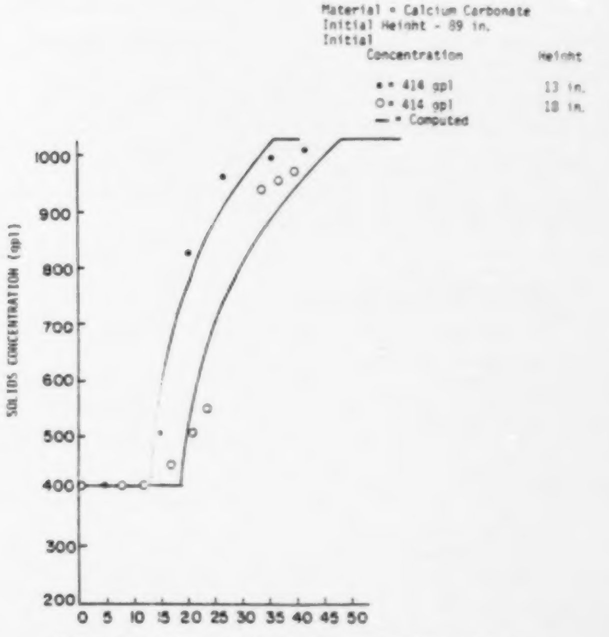

Fig. 6. Batch concentration profiles for calcium carbonate.

TABLE 1. PREDICTED SOLIDS CONCENTRATIONS AT WHICH SIGNIFICANT DEVIATION FROM FITTED AND MEASURED HEIGHTS IN SETTLING CURVES BEGINS (REFER TO FIGURE 4)

Initial concentration (g/l)	C_s (g/l)	Actual height (in.)	Predicted height (in.)	Time (hr)
235	712	22.5	23.5	30
300	857	29.5	30.7	33
413	883	37	38.3	51
440	924	37	38.2	70
AVG. = 844				

Comparisons of predicted vs. measured concentration profiles existing across the sludge bed at two different settling times and of predicted vs. measured concentration variations with time at two different heights above the bottom are given in Figures 5 and 6, respectively. Below predicted solids concentrations C_s of about 844 g/l, the theoretical and actual concentration profiles agree well. For $C_s > 844$ g/l, however, not only the theoretical and actual concentration profiles but also the fitted and actual settling curves (Table 1) deviate significantly. Apparently, then, Kynch's analysis and thus the proposed procedure is valid within dilute concentration ranges (for example, for this case $C_s < 844$ g/l).

In order to verify the model for continuous settling, described in the next section, it was necessary to have both batch and continuous data collected on the same sludge. Only one set of data was found which could be used for this purpose, and that was collected by Dorr-Oliver (*4*) utilizing a calcium carbonate slurry; similar data are now being collected at Auburn for other sludge systems.

Using the Dorr-Oliver (*4*) data, six sets of batch tests were fitted simultaneously in order to determine the four parameters which were used to predict the concentration profiles shown in a later section. Figure 7 is presented as a typical batch plot for the Dorr-Oliver (*4*) data and illustrates the type of fit obtained. Only at the highest concentration was there

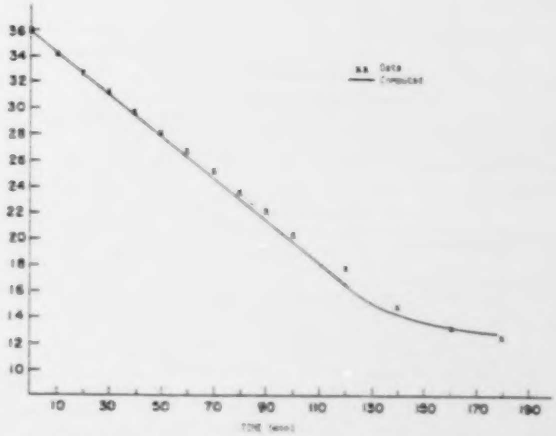

Fig. 7. Batch interface subsidence curve for calcium carbonate (*4*).

significant deviation of the model from the actual data, indicating again the validity of the batch model. Although as good a representation as possible is desired for the batch data, it should be kept in mind that emphasis should be placed on the agreement obtained for the continuous system.

DEVELOPMENT OF THE MODEL

An approach different from that typically taken, yet consistent with standard design practice, was assumed in modeling the continuous process. As is usual practice, the total solids flux N_s was assumed to be the sum of the fluxes due to subsidence of the particles vC_s (that is, gravity settling) and the flux due to the bulk downward movement of sludge $C_s\Psi$

$$N_s = C_s(\Psi + v) \tag{8}$$

Yet, Ψ was not defined as the bulk downward volumetric flow of sludge, as is typically done (Table 2). Instead, Equation (8) was taken as the defining equation for Ψ since, in consistency with the characterization procedure (Kynch theory) and standard design procedures, $v = f(C_s)$ and thus Ψ is determined:

$$\Psi \equiv N_s/C_s - v \tag{9}$$

If, on the other hand, Ψ were defined to be the bulk downward volumetric (or mass) velocity u_v, then v would become determined:

$$v = N_s/C_s - u_v \tag{10}$$

In addition, since u_v is a constant throughout the sludge blanket (*13*), v must also be a constant. C_s is known to vary throughout the sludge blanket [Figure 8, (*1*)]; thus, $v \neq f(C_s)$, which is a direct violation of Kynch theory and experimental

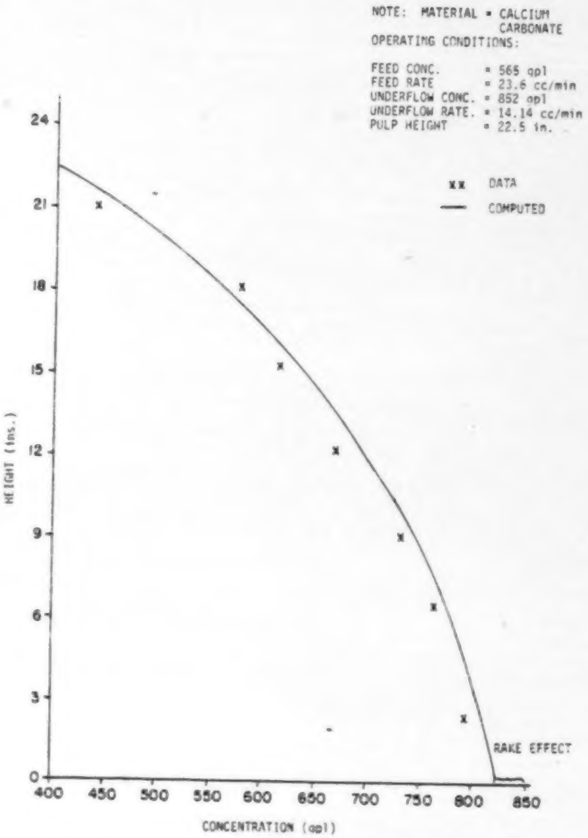

Fig. 8. Concentration profile for continuous thickener.

TABLE 2. DERIVATION OF RELATIONSHIP BETWEEN SETTLING VELOCITY AND SOLIDS CONCENTRATION AND BETWEEN BULK VELOCITY DEFINED ON BASIS OF VOLUMETRIC FLOW AND BULK VELOCITY AS USED IN THE PAPER (*13*)

Volumetric flow	As defined in paper
Solids: $v_s f = \frac{N_s}{\rho_s}$ = constant Liquid: $v_1(1-f) = \frac{N_1}{\rho_1}$ = constant Total: $u_v = v_s f + v_1(1-f)$ Setting: $u = u_v$ $v_s f = \frac{c_s}{\rho_s}(u_v + v) = \frac{N_s}{\rho_s}$ $v = \frac{N_s}{C_s} - u_v$ This determines $v(c_s)$, since N_s and u_v are both constants; as a result, v is not free to be specified from batch data	$N_s = C_s(\Psi + v)$ = constant $v = f(c_s)$, empirical relationship determined from batch data Thus, once c_s is specified $\Psi \equiv (N_s/c_s) - v$ Then c_1, w, ρ, u^*, and u_v can be determined: $N_1 = c_1(\Psi + w)$ = constant $c_1 = \rho_1 \frac{(1 - c_s)}{\rho_s}$ $w = (N_1/c_1) - \Psi$ $\rho = c_s + c_1$ $N_T = \rho_{st}U_v$ = constant $N_T = c_s(\Psi + v) + c_1(\Psi + w)$ = constant $f = \frac{c_s}{c_1}$; $v_s = \frac{c_1}{\rho_s}(\Psi + v)$ $v_1 = \frac{c_s}{\rho_1}(\Psi + w)$ $\rho_{st} = \frac{\rho_s v_s f + \rho_1 v_1(1-f)}{v_s f + v_1(1-f)}$ $u_v = N_T/\rho_{st}$

observation as determined, that is, in a batch settling system. By taking Equation (9) as the defining equation for Ψ, not forcing it to be a constant (that is, $\Psi = u_v$), as is done here, no direct violation of experimental observation [that is, $v = f(C_s)$] results. Actually, $\Psi + v = u_v + v^*$ where v^* is used to denote v as determined by Equation (10), and the frame of reference is the same for each; that is, the net downward velocity of solids $= \Psi + v = u_v + v^*$ and

$$\Delta Z = \int_0^{\Delta t} (\Psi + v)\,dt = \int_0^{\Delta t} (u_v + v^*)\,dt \qquad (11)$$

where ΔZ is the distance moved by the subsiding solids in time Δt. Furthermore, it should be noted that in the limit as v becomes insignificant relative to Ψ (that is, $v <<< \Psi$), $\Psi = u_v$; conversely, for $\Psi = 0$ (batch settling), v becomes equal, by definition, to the velocity of subsiding solids in the sludge blanket relative to the vessel walls.

To predict the variation of Ψ, v, and C_s across the sludge blanket, an assumption must be made with regard to their dependence on residence time. As observed by Comings (*1*), the concentration of a layer of sludge varies continuously throughout the sludge blanket in a manner similar to that occurring in the hindered settling phase of batch settling; Figures 5, 8, and 9 confirm this observation. If we assume, therefore, that this observation is the case, then a hypothetical batch settling system should exist for which the change in C_s with time during hindered settling is the same as that in the sludge blanket. The Talmadge-Fitch equation, which gives $C_s = f(t)$ in a batch system, was modified so as to represent thickening within this hypothetical batch system

$$C_s = C_{ss} Z_{1B}/(Z_B + vt) \qquad (12)$$

where C_{ss} is the initial concentration. Z_B is the interface height in the hypothetical batch system

$$Z_B = \int_t^{tu} v\,dt + Z_{Bu} \qquad (13)$$

Z_{Bu} is the interface height at which underflow concentration C_u exists:

$$Z_{Bu} = -t_u d(Cv)/dC\big|_{C_u} \qquad (14)$$

Z_{1B} is the initial height of the hypothetical batch system

$$Z_{1B} = C_u(Z_{Bu} + t_u v\big|_{C_u})/C_{ss}$$

or

$$= -t_u C^2 \Psi(dv/dC)\big|_{C_u}/C_{ss} \qquad (15)$$

Once C and v have been computed using Equations (12), (13), (14), and (15), Ψ can be determined from Equation (9), and the height of the sludge blanket Z_H necessary to attain the desired underflow concentration C_u is given by

$$Z_H = \int_{t_D}^{t_u} (v + \Psi)\,dt \qquad (16)$$

where t_D is the time at which hindered settling begins in the associated batch process.

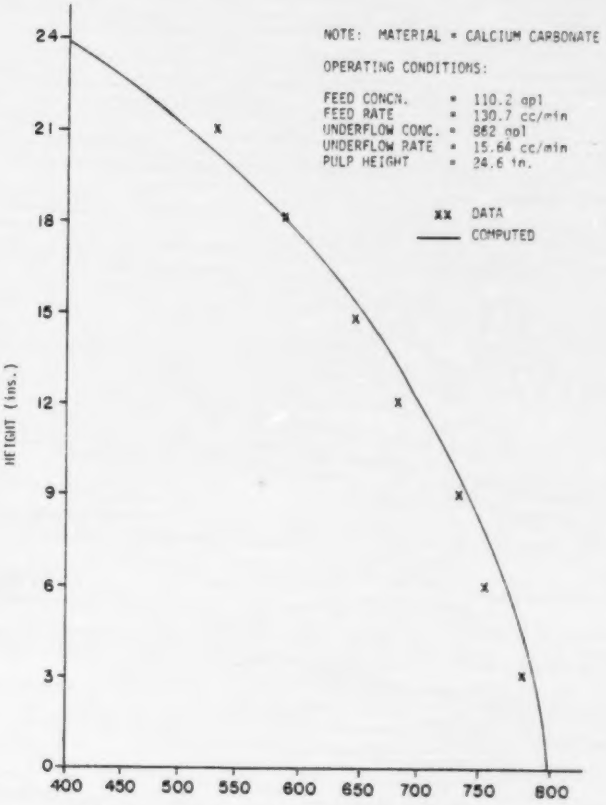

Fig. 9. Concentration profile for continuous thickener.

In the previous development, the concentration C_{ss} in the settling zone is approximated as being constant. For calculation of C_{ss}, a correlation relating it with feed rate, feed mechanism (central or peripheral), and so forth may be determined. Or, as done in this work, the bulk velocity (Ψ) in the settling zone may be estimated as a multiple R of the underflow velocity. This assertion can be used in turn to estimate C_{ss} as follows:

$$C_{ss} = G/(\Psi + v\big|_{C_{ss}})$$
$$= C_u \cdot U_u/[R \cdot U_u + B_o \exp(B_1 C_s + B_2 C_s^2) + B_3] \qquad (17)$$

where U_u is the bulk velocity at the underflow.
At the underflow, practically little settling is occurring; thus v is taken to be zero, and $\Psi = u_v$.

EVALUATION OF THE MODEL

From the parameter values obtained for the calcium carbonate data of Dorr-Oliver (*4*), nine concentration profiles were computed for different inlet flows and concentrations and different underflow concentrations. Figures 9 and 10 show rep-

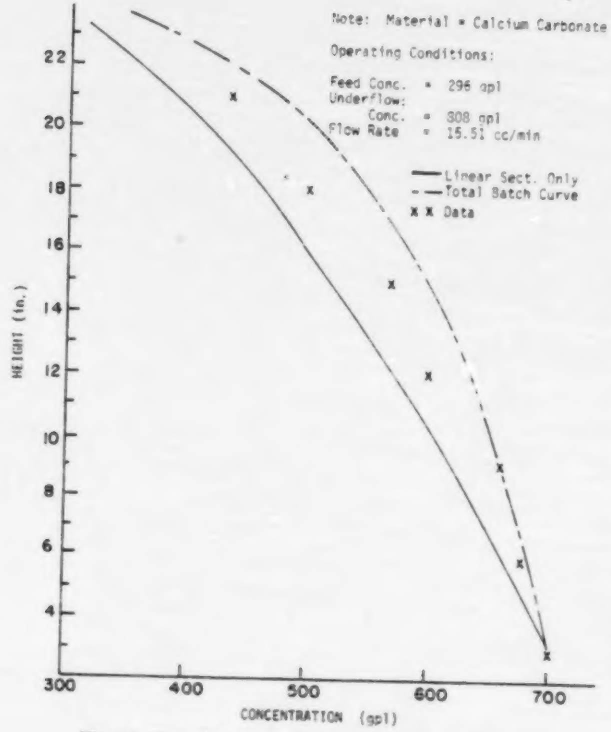

Fig. 10. Sensitivity with respect to parameter values.

resentative plots which are typical of the fits obtained. In general, the fits are quite satisfying, indicating that the proposed model is capable of predicting concentration profiles from batch settling data. There are, however, several points which should be emphasized. With reference to Figure 9, it can be seen that a very rapid increase in concentration occurs in the bottom of the continuous settler. This occurred in all of the continuous data and is thought to be the result of the raking action in the bottom of the settler similar to that observed by Comings (*1*). Because the magnitude of this effect will depend upon the geometric configuration of the rake, no effort was made to model it. In actual practice this could be incorporated as a safety factor.

Estimation of C_{ss} represents a definite weakness in the continuous model. As previously discussed, the bulk velocity in the settling zone was set equal to a multiple of the underflow velocity. This approach usually yielded better results than those obtained with another approximation based on a water balance. An interesting observation was that normally the feed stream is concentrated upon entry into the thickening section. Dick (*2*) has postulated that dilution of the feed stream occurs, but this was found to happen only at high feed concentrations.

Some controversy exists over which portion of the batch settling curve should be employed to determine the parameters relating settling velocity to solids concentration. Figure 10 shows the predictions obtained when the parameters come only from the linear portion of the curve or from the entire curve weighted equally. A compromise between these two cases seems to be most promising.

In addition to facilitating analysis of thickener operation (predicting concentration profiles, etc.), the model can also be effective as a design tool. Given a maximum allowable residence time and a desired total solids underflow ($A \cdot N_s$), an optimum settler design could be arrived at by balancing area, height, and underflow concentration so as to minimize cost. If the case is such that both height and residence time of solids are fixed, only area or underflow concentration could be manipulated so as to minimize cost while the other would be set in order to satisfy the constraints placed by the proposed model. To illustrate the effectiveness of the model in this capacity, operating conditions for a 1 ft diameter pilot plant unit (*4*) were used to predict area requirement, followed by comparing the predicted value with the actual area. For each set of operating conditions considered, the underflow concentration and the underflow rate were known. The mean residence time was estimated from known concentration profile data and used as a basis for estimating the area requirement. Table 3 lists some values computed in this manner along with those predicted by the method advocated by Dick (*2, 3*) as well as the actual area. The settling rate expression upon which these computations were based was determined, while data contained within both free and hindered settling regions of several batch settling tests were considered as previously described. If only the free settling region of each batch test had been considered, the area requirements predicted by Dick's procedure would have been lower. The area requirements predicted while using the proposed model agree with the actual value well within engineering accuracy and serve to demonstrate its usefulness in design applications.

CONCLUSIONS

The degree of thickening achieved in a sedimentation basin is influenced greatly by the residence time of solids and the

TABLE 3. COMPARISON OF ACTUAL AREA REQUIREMENTS AND PREDICTED VALUES

Feed concn, g/l	Underflow concn,g/l	Underflow rate, cm^3/min	Predicted area (model), cm^2	Predicted area (Dick's procedure), cm^2	Actual area cm^2
109.2	725	19.95	739	890	730
172.0	621	36.3	741	857	730
499	766	15.08	739	909	730

The settling rate expression upon which both area computations were based was determined while considering data contained within both free and hindered settling regions of several batch settling test (*4*).

height of the sludge blanket formed at the bottom of the basin. A simple method has been presented which allows consideration of these factors, as well as those commonly treated, while determining the thickening capacity of a continuous system. The technique is applicable only to systems operating at relatively low underflow concentrations, where compression does not occur to an appreciable extent. The effectiveness of the method has been demonstrated by comparing predicted concentration profiles with corresponding experimental data for calcium carbonate, showing excellent agreement.

ACKNOWLEDGMENT

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NOTATION

A = area
B_k = constant parameters, Equation (6)
C_1 = liquid concentration
C_s = suspended solids (SS) concentration
C_{sf} = SS concentration at which batch subsidence ceases
C_{ss} = SS concentration in the settling zone (continuous system)
f = fraction of area through which solids pass
H = height from bottom of column
N_1, N_s, N_t = liquid, solid, and total flux, respectively
t = time
t_k = time at which concentration gradient reaches interface
U_u = bulk velocity underflow
u_v = bulk volumetric flow
v = settling velocity
w = rate of upward propagation toward sludge-supernatant interface
z = height from bottom of column
Z_{1B} = initial height in hypothetical batch system, Equation (13)
Z_B = height in hypothetical batch system, Equation (15)
Z_H = height of sludge blanket
ρ = density
Ψ = bulk velocity, Equation (9)

SUBSCRIPTS

i, o = initial values
1 = value for liquid fraction
s = value for solid fraction
u = value of variable in underflow

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Appendix B

Filtration Notation and Computational Procedures

1. Filtration Notation
2. Sample Calculation

FILTRATION NOTATION

I. Flow through a packed bed whose thickness is changing with time. Slurry is separated into cake and filtrate.

II. Flow is almost always streamline, so we can apply Poiseuille's Equation for flow through a passage:

$$-\frac{dp}{dL'} = \frac{32u_b\mu}{g_cD^2}$$

p = pressure
L' = passage length
u_b = average velocity in passage
μ = liquid (filtrate) viscosity
D = passage diameter

In order to apply this equation, we must relate u_b to the filtrate flow rate, and express L' and D in terms of cake properties.

III. Let

V = filtrate volume
θ = time
A' = flow area through cake
ε = cake porosity
A = Superficial flow area (cake cross-sectional area)
$\frac{dV}{d\theta}$ = volumetric flow rate of filtrate

$= u_bA' = u_bA\varepsilon$

IV. Since our passages are not circular, we replace D by the equivalent diameter, Deq, which is equal to 4 times the hydraulic radius, r_H

$$r_H = \frac{\text{cross-sectional area of passage}}{\text{wetted perimeter of passage}}$$

Multiplying numerator and denominator by the number of passages and the length of passages,

$$r_H = \frac{\text{void volume of cake}}{\text{exposed surface area of particles}}$$

Assuming that the area of contact of the particles is a negligible portion of the total surface area of the particles,

$$r_H = \frac{\varepsilon \text{ x cake volume}}{Sv(1-\varepsilon) \text{ x cake volume}}$$

where

S_v = surface area/volume for the particles

then

$$\frac{-dV}{d\theta} = \frac{g_c dp}{\mu dL'} A \text{ x } \frac{\varepsilon^2}{2(1-\varepsilon)^2 Sv^2}$$

V. We do not know the length of our passages, and, in fact, they may vary widely. L´, then must represent some statistical average (as does S_v). We relate it to the cake thickness by defining a proportionally constant

$$L' = \frac{k}{2} L$$

where

L = cake thickness

For spherical particles k is knwon to be 5. 4.17 may be a more generally valid figure for non-spherical particles.

$$\frac{-dV}{d\theta} = \frac{g_c dp}{\mu dL} A \frac{\varepsilon^3}{k(1-\varepsilon)^2 Sv^2}$$

VI. In order to proceed, it is necessary to eliminate the derivative, $\frac{dp}{dL}$.

This is done by writing the equation as

$$-\int_{p_i}^{P} \frac{dp}{\frac{k(1-\varepsilon)Sv^2}{\rho_s \varepsilon^3}} = \int_{L}^{0} \left[\frac{\mu \rho_s (1-\varepsilon)}{A g_c} \frac{dV}{d\theta}\right] dL$$

and integrating over the cake thickness

$$\frac{P-p_i}{\alpha_{Av}} = \frac{\mu\rho_s}{Ag_c}\frac{dV}{d\theta}\; L(1-\varepsilon_{Av.})$$

It should be noted that $\alpha_{Av.}$ depends on P, p_i, and the functional dependence of α on p. The primary factor in this dependence is ε.

also

$$\varepsilon_{Av.} = 1 - \frac{\int_0^L (1-\varepsilon)dL}{\int_0^L dL} = 1 - \frac{\int_0^L (1-\varepsilon)dL}{L} = \frac{\int_0^L \varepsilon dL}{L}$$

Thus

$$\frac{P - p_i}{L(1-\varepsilon_{Av.})} = \alpha_{Av.}\frac{\mu\rho_s}{Ag_c}\frac{dV}{d\theta} = \frac{\Delta P_c}{L(1-\varepsilon_{Av.})}$$

where

ΔP_c = pressure drop across cake

ρ_s = particle densities

VII. It turns out to be convenient to define another quantity, L_m, such that

$$\frac{\Delta P}{L+L_m} = \frac{\Delta P_c}{L}$$

Where L_m is the thickness of cake necessary to produce all additional pressure drops (beside that across the cake) between those points across which ΔP is measured. This is necessary, because ΔP_c cannot usually be measured. We then have

$$\Delta P = \alpha_{Av.}\frac{\mu\rho_s}{Ag_c}\frac{dV}{d\theta}\,(L+L_m)\,(1-\varepsilon_{Av.})$$

VIII. This equation is still quite general, and is, in fact, applicable to flow in packed beds of all kinds, as long as the flow is viscous. In the particular application to filtration, we make use of the fact that the cake thickness is related to the filtrate volume. Thus:

$$\text{wt. of cake solid} = LA\rho_S\,(1-\varepsilon_{Av.}) = (V+LA\varepsilon_{Av.})\rho\,\frac{w}{1-w}$$

where

w = wt. fraction of solids in slurry

solving

$$L = \frac{V}{A}\,\frac{\rho w}{\rho_S(1-\varepsilon_{Av.})(1-w)-\varepsilon_{Av.}\rho w}$$

and similarly L_m can be expressed in terms of V_m. Then, we have

$$\Delta P = \frac{\alpha_{Av}\mu\rho_S\rho w(1-\varepsilon_{Av})}{A^2 g_c[\rho_S(1-\varepsilon_{Av})(1-w)-\varepsilon_{Av}\rho w]}\;\frac{dV}{d\theta}\,(V+V_m)$$

or

$$\Delta P = \frac{\alpha_{Av}\mu\rho_S(1-\varepsilon_{Av})}{g_c\rho_w}\,[\rho_S(1-\varepsilon_{Av})(1-w)-\varepsilon_{Av}\rho w]\,\frac{dL}{d\theta}\,(L+L_m)$$

IX. For plate and frame type filters, it is more convenient to work in terms of filtrate volume, so that we rewrite the volume equation as

$$\frac{dV}{d\theta}\,(V+V_m) = \frac{A^2\Delta P}{2C_v}$$

where

$$C_v = \frac{\alpha_{Av}\mu\rho_S\rho w(1-\varepsilon_{Av})}{2g_c[\rho_S(1-\varepsilon_{Av})(1-w)-\varepsilon_{Av}\rho w]}$$

or if α is constant (incompressible cake)

$$C_v = \frac{kS_v^2(1-\varepsilon)^2\mu\rho w}{2g_c\varepsilon^3[\rho_s(1-\varepsilon)(1-w)-\varepsilon\rho w]}$$

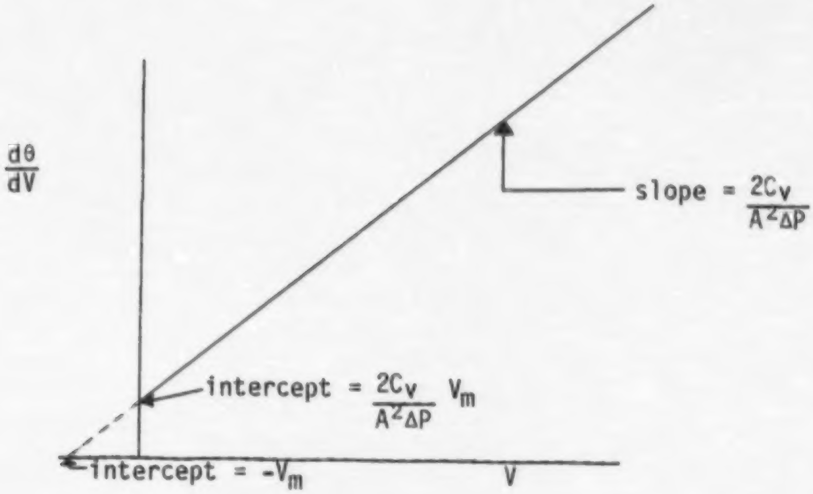
X. Incompressible Cakes (C_v = constant)

A. Constant Pressure Filtration (ΔP = const.)

1. Determination of Parameters

$$\frac{dV}{d\theta} = \frac{A^2\Delta P}{2C_v(V+V_m)}$$

$$\frac{d\theta}{dV} = \frac{2C_v}{A^2\Delta P}\ (V + V_m)$$



Experimentally, a plot of $\frac{\Delta\theta}{\Delta V}$ vs $V_{Av.}$, is frequently easier to prepare.

2. Prediction of Performance

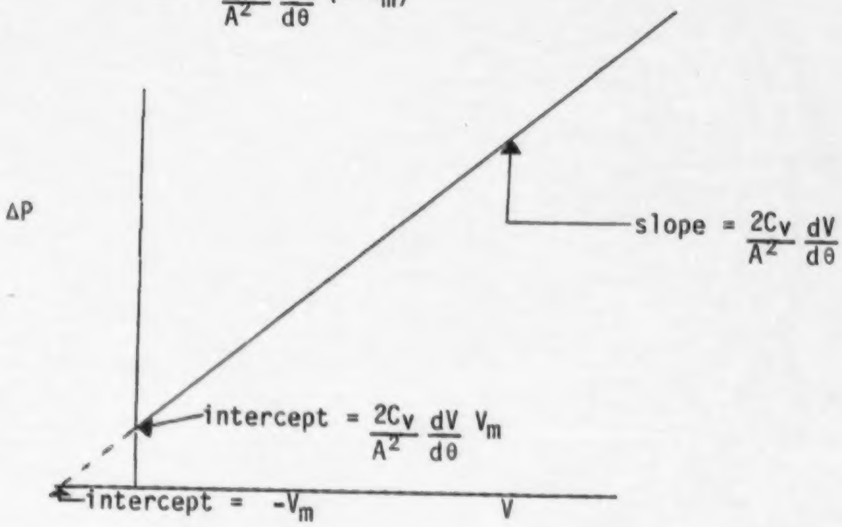
$$\int_0^\theta d\theta = \frac{2C_v}{A^2\Delta P}\int_0^V (V+Vm)\,dV$$

$$\theta = \frac{C_v}{A^2\Delta P}V^2 + \frac{2C_v}{A^2\Delta P}V_mV$$

B. Constant Rate Filtration, $\frac{dV}{d\theta}$ = const.

1. Determination of Parameters

$$\Delta P = \frac{2C_v}{A^2}\frac{dV}{d\theta}(V+V_m)$$



2. Prediction of Performance

$$dV = \frac{dV}{d\theta}d\theta$$

$$V = \frac{dV}{d\theta}\theta$$

$$\Delta P = \frac{2C_v}{A^2} \frac{V}{\theta} (V+V_m)$$

$$\Delta P = \frac{2C_v}{A^2} \frac{V^2}{\theta} + \frac{VV_m}{\theta}$$

XI. Compressible Cakes

A. Constant Pressure Filtration (ΔP = const.)

For this case, $\alpha_{Av.}$ is constant during the filtration, but $\varepsilon_{Av.}$ is not. Also, data taken at one pressure drop cannot be used at another. A frequent assumption used to obtain $\alpha_{Av.}$ is to use

$$\alpha = \beta p^r$$

But results are very hard to reproduce and $\varepsilon_{Av.}$ is still a problem.

B. Constant Rate Filtration

$\alpha_{Av.}$ and $\varepsilon_{Av.}$ are both varying during the filtration.

SAMPLE CALCULATIONS

(A.) CONVERSION OF FILTRATE VOLUME: V

For run 1, $V = 1\ell$.

$$V(\text{ft}^3) = \frac{1\ell}{}\,\Big|\,\frac{1000\text{ cm}^3}{1\ell}\,\Big|\,\frac{1\text{m}^3}{(100)^3\text{cm}^3}\,\Big|\,\frac{35.31\text{ft}^3}{1\text{m}^3}$$

$$V = 0.03531\text{ ft}^3$$

(B.) CONVERSION OF PRESSURE: P

For run 1, $P = 5.5$ psig.

$$P(\text{lb/ft}^2) = \frac{5.5\text{ lb}}{\text{in}^2}\,\Big|\,\frac{144\text{ in}^2}{\text{ft}^2}$$

$$P = 792\text{ lb/ft}^2$$

(C.) CALCULATION OF WEIGHTS OF FILTER CAKE:

For Run 1,

1. $\text{Wt. of Wet Cake} = \frac{(1712.7-508.5)\text{g}}{}\,\Big|\,\frac{\text{lb}}{454\text{g}} = 2.6524\text{ lb.}$

2. $\text{Wt. of Dry Cake} = \frac{(808.2 - 508.5}{}\,\Big|\,\frac{\text{lb}}{454\text{g}} = 0.6601\text{ lb.}$

(D.) CONVERSION OF PARTICLE DENSITY: ρ_s

$$\rho_s\ (\text{lb/ft}^3) = \frac{1.65\text{g}}{\text{cm}^3}\,\Big|\,\frac{\text{lb}}{454\text{g}}\,\Big|\,\frac{(100)^3\text{cm}^3}{1\text{m}^3}\,\Big|\,\frac{1\text{m}^3}{35.31\text{ft}^3}$$

$$\rho_s = 102.93\text{ lb/ft}^3$$

(E.) CALCULATION OF VOLUMES OF FILTER CAKE:

For Run 1,

1. Volume of Dry Cake $= \begin{array}{c|c} 0.6601\ \text{lb} & \text{ft}^3 \\ \hline & 102.93\ \text{lb} \end{array}$

$$= (6.413 \times 10^{-3})\text{ft}^3$$

2. Void Volume = Volume of H_2O Evaporated

$$= \begin{array}{c|c} (2.6524-0.6601)\text{lb} & \text{ft}^3 \\ \hline & 62.26\ \text{lb} \end{array}$$

$$= 0.03200\ \text{ft}^3$$

3. Volume of Wet Filter Cake $= ((6.413 \times 10^{-3})+0.03200)\ \text{ft}^3$

Total Volume $= 0.03841\ \text{ft}^3$

(F.) CALCULATION OF POROSITY: ε

For Run 1,

$$\varepsilon = \frac{\text{Void Volume}}{\text{Total Cake Volume}} = \frac{0.03200\ \text{ft}^3}{0.03841\ \text{ft}^3}$$

$$\varepsilon = 0.8331$$

(G.) CALCULATION OF WEIGHT FRACTION OF THE SLURRY SOLIDS: w

For Run 1,

$$w = \frac{\text{wt. of dry cake}}{(\text{wt. of wetcake}) + (\text{wt. of filtrate})}$$

$$w = \frac{0.6601\ \text{lb}}{(2.6524\ \text{lb}) + \left(\begin{array}{c|c} 0.91806\ \text{ft}^3 & 62.26\text{lb} \\ \hline & \text{ft}^3 \end{array}\right)}$$

$$w = 0.01104$$

(H.) CALCULATION OF S_v:

$$C_v = \frac{kS_v^2(1-\varepsilon)^2 \ \mu\rho w}{2gc \ \varepsilon^3 \ [\rho_s(1-\varepsilon)(1-w)-\varepsilon\rho w]}$$

or

$$S_v = \left[\frac{2gc \ \varepsilon^3 C_v \ [\rho_s(1-\varepsilon)(1-w)-\varepsilon\rho w]}{k(1-\varepsilon)^2\mu\rho w}\right]^{1/2}$$

For Run 1,

g_c = gravitational constant = $32.2 \ \frac{\text{lbm} \cdot \text{ft}}{\text{lb}_f \cdot \text{sec}^2}$

ε = porosity = 0.8331

k = 4.17

C_v = (from graph) = $1.392 \times 10^6 \ \frac{\text{lb.sec}}{\text{ft}^4}$

ρ_s = sample density = 102.93

ρ (@23.5°C) = filtrate (assume H_2O) density = 62.26 lb/ft^3

w = weight fraction of slurry solids = 0.01104

μ = viscosity filtrate = $1.9245^{-5} \text{lb}_f \ \text{sec/ft}^2$

= 6.1969×10^{-4} lbm/ft·sec

$$S_v = \left[2 \left| \frac{32.2\text{lbmft}}{\text{ft}_f \cdot \text{sec}^2} \right| (0.8331)^3 \left| \frac{1.392\text{x}10^6\text{lb.sec}}{\text{ft*}^2} \right| \frac{}{4.17} \left| \frac{}{(1.0-0.8331)^2} \right| \frac{\text{ft} \cdot \text{sec}}{6.1969\text{x}10^{-4}\text{lb}_m}\right.$$

$$\left. \times \frac{\text{ft}^3}{62.26 \ \text{lb}} \left| \frac{}{0.01104} \right. \left[\frac{102.93\text{lb}}{\text{ft}^3} \Big| \underline{(1-0.8331)(1-0.01104)} - \underline{0.8331 \Big| \frac{62.26\text{lb}}{\text{ft}^3} \Big| 0.01104}\right]\right]^{1/2}$$

$S_v = 4.1472 \times 10^6 \ \text{ft}^{-1}$

(I.) CALCULATION OF THE SURFACE AREA OF THE FILTER CAKE:

$A = (S_v)(\text{Volume of Dry Cake})$

For Run 1,

$$A = \frac{4.1472 \times 10^6}{\text{ft}} \Bigg| (6.413 \times 10^{-3})\ \text{ft}^3$$

$A = 2.6596 \times 10^4\ \text{ft}^2$

Appendix C

Details of Economic Study Performed By Engineering Staff of Tennessee Valley Authority

TABLE I

Base Case[a]

Total Capital Investment - Process Equipment and Installation Analysis

	Total, k$	Percent of direct investment	Percent of total capital investment
Process equipment	2,551	52.2	23.7
Piping and insulation	176	3.6	1.6
Foundation and structural	136	2.8	1.3
Excavation, site preparation, roads and railroads	136	2.4	1.1
Electrical	639	13.1	6.0
Instrumentation	70	1.4	0.7
Buildings	550	11.3	5.1
Subtotal	4,237	86.8	39.5
Services and miscellaneous	64	1.3	2.6
Subtotal excluding trucks and equipment	4,301	88.1	40.1
Trucks and earthmoving equipment	581	11.9	5.4
Subtotal direct investment	4,882	100.0	45.5
Engineering design and supervision	392	8.0	3.7
Architect and engineering contractor	98	2.0	0.9
Construction expense	839	17.2	7.8
Contractor fees	320	6.6	3.0
Subtotal	6,531	133.8	60.9
Contingency	1,306	26.7	12.2
Subtotal fixed investment	7,837	160.5	73.1
Allowance for startup and modifications	726	14.9	6.8
Interest during construction	940	19.3	8.8
Subtotal capital investment	9,503	194.7	88.7
Land	676	13.8	6.3
Working capital	538	11.0	5.0
Total capital investment	10,717	219.5	100.0

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[a] Basis

IUCS process base case with conventional dewatering system.
New 500-MW plant (30-yr life); 198 klb/hr (60% solids) sludge.
Midwest plant location; average basis for cost scaling, mid-1979.
Coal analysis (by weight): 3.5% sulfur (dry basis), 16% ash.
Fly ash removed with SO_2.
SO_2 and fly ash removed to meet NSPS.
Limestone process with 1.5 stoichiometry based on SO_2 removed.
Disposal in landfill 1 mile from scrubber facilities, 60% solids cake.
Land required for disposal, 193 acres at 30 ft deep.
Trucks used for transport of treated sludge to disposal area.
Lime added at 4% of dry solids.

TABLE II

Proposed System (Modified Dewatering; Case I)[a]

Total Capital Investment - Process Equipment and Installation Analysis

	Total, k$	Percent of direct investment	Percent of total capital investment
Process equipment	2,240	51.5	23.2
Piping and insulation	154	3.5	1.6
Foundation and structural	120	2.8	1.2
Excavation, site preparation, roads and railroads	103	2.4	1.1
Electrical	562	12.9	5.8
Instrumentation	60	1.4	0.6
Buildings	485	11.1	5.0
Subtotal	3,724	85.6	38.5
Services and miscellaneous	47	1.1	0.5
Subtotal excluding trucks and equipment	3,771	86.7	39.0
Trucks and earthmoving equipment	581	13.3	6.0
Subtotal direct investment	4,352	100.0	45.0
Engineering design and supervision	403	9.3	4.2
Architect and engineering contractor	101	2.3	1.0
Construction expense	752	17.3	7.8
Contractor fees	294	6.8	3.0
Subtotal	5,902	135.7	61.0
Contingency	1,180	27.1	12.2
Subtotal fixed investment	7,082	162.8	73.2
Allowance for startup and modifications	656	15.1	6.8
Interest during construction	851	19.6	8.8
Subtotal capital investment	8,589	197.5	88.8
Land	676	15.5	7.0
Working capital	404	9.3	4.2
Total capital investment	9,669	222.3	100.0

---

[a] Basis

Case I IUCS process with modified dewatering system. The thickener in the conventional dewatering train has been replaced by the proposed dewatering system (Auburn study).

New 500-MW plant (30-yr life); 198 klb/hr (60% solids) sludge.

Midwest plant location; average basis for cost scaling, mid-1979.
Coal analysis (by weight): 3.5% sulfur (dry basis), 16% ash.
Fly ash removed with SO_2.
SO_2 and fly ash removed to meet NSPS.
Limestone process with 1.5 stoichiometry based on SO_2 removed.
Disposal in landfill 1 mile from scrubber facilities, 60% solids cake.
Land required for disposal, 193 acres at 30 ft deep.
Trucks used for transport of treated sludge to disposal area.
Lime added at 4% of dry solids.

TABLE III

Proposed System (Modified Dewatering; Case II)[a]

Total Capital Investment - Process Equipment and Installation Analysis

	Total, k$	Percent of direct investment	Percent of total capital investment
Process equipment	2,329	51.8	24.0
Piping and insulation	161	3.6	1.7
Foundation and structural	124	2.7	1.3
Excavation, site preparation, roads and railroads	107	2.4	1.1
Electrical	584	13.0	6.0
Instrumentation	62	1.4	0.6
Buildings	504	11.2	5.2
Subtotal	3,871	86.1	39.9
Services and miscellaneous	43	1.0	0.4
Subtotal excluding trucks and equipment	3,914	87.1	40.3
Trucks and earthmoving equipment	581	12.9	6.0
Subtotal direct investment	4,495	100.0	46.3
Engineering design and supervision	392	8.7	4.0
Architect and engineering contractor	98	2.2	1.0
Construction expense	776	17.3	8.0
Contractor fees	301	6.7	3.1
Subtotal	6,062	134.9	62.4
Contingency	1,114	24.8	11.5
Subtotal fixed investment	7,176	159.7	73.9
Allowance for startup and modifications	674	15.0	6.9
Interest during construction	804	17.9	8.3
Subtotal capital investment	8,654	192.6	89.1
Land	676	15.0	7.0
Working capital	383	8.5	3.9
Total capital investment	9,713	216.1	100.0

---

[a] Basis

Case II IUCS process with modified dewatering system. The thickener and filter in the conventional dewatering train have been replaced by the proposed dewatering system (Auburn study).
New 500-MW plant (30-yr life); 198 klb/hr (60% solids) sludge.
Midwest plant location; average basis for cost scaling, mid-1979.
Coal analysis (by weight): 3.5% sulfur (dry basis), 16% ash.
Fly ash removed with SO_2.
SO_2 and fly ash removed to meet NSPS.
Limestone process with 1.5 stoichiometry based on SO_2 removed.
Disposal in landfill 1 mile from scrubber facilities, 60% solids cake.
Land required for disposal, 193 acres at 30 ft deep.
Trucks used for transport of treated sludge to disposal area.
Lime added at 4% of dry solids.

TABLE IV

Base Case[a]

Total Annual Revenue Requirements - Regulated Utility Economics

	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Direct costs				
Delivered raw materials				
Lime (ground)	16,216 tons	53.00/ton	859,400	16.2
Subtotal raw materials cost			859,400	16.2
Conversion costs				
Operating labor and supervision				
Plant	35,040 man-hr	12.50/man-hr	438,000	8.3
Solids disposal equipment	52,560 man-hr	17.00/man-hr	893,500	16.9
Maintenance--plant labor and material, 4% of direct investment			172,000	3.3
Landfill operation				
Land preparation			11,000	0.2
Trucks (fuel and maintenance)	693,000 tons	0.06/ton	41,600	0.8
Earthmoving equipment (fuel and maintenance)	693,000 tons	0.16/ton	110,900	2.1
Electricity	3,688 kkWh	0.029/kWh	107,000	2.0
Analyses	1,000 hr	17.00/man-hr	17,000	0.3
Subtotal conversion costs			1,791,000	33.0
Subtotal direct costs			2,650,400	50.1
Indirect costs				
Capital charges				
Depreciation, interim replacement, and insurance at 7.83% of total capital investment less land and working capital			744,100	14.1
Average cost of capital and taxes at 8.6% of total capital investment			921,700	17.4
Overhead				
Plant, 50% of conversion costs less utilities			842,000	15.9
Administrative, 10% of operating labor			133,200	2.5
Subtotal indirect costs			2,641,000	49.9
Total annual revenue requirements			5,291,400	100.0

	$/dry ton	$/wet ton	Mills/kWh
Equivalent unit revenue requirements	12.55	7.63	1.51

---

a. Basis

IUCS process with conventional dewatering system.
Remaining plant life, 30 yr.
Coal burned, 482,571 lb/hr, 9,000 Btu/kWh.
Power plant on-stream time, 7,000 hr/yr.
Total capital investment, $10,717,000.
Midwest plant location, mid-1980 operating costs.

TABLE V

Proposed System (Modified Dewatering; Case I)[a]

Total Annual Revenue Requirements - Regulated Utility Economics

	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Direct costs				
Delivered raw materials				
Lime (ground)	16,216 tons	53.00/ton	859,400	16.7
Subtotal raw materials cost			859,400	16.7
Conversion costs				
Operating labor and supervision				
Plant	35,040 man-hr	12.50/man-hr	438,000	8.5
Solids disposal equipment	52,560 man-hr	17.00/man-hr	893,500	17.4
Maintenance—plant labor and material, 4% of direct investment			174,100	3.4
Landfill operation				
Land preparation			11,000	0.2
Trucks (fuel and maintenance)	693,000 tons	0.06/ton	41,600	0.8
Earthmoving equipment (fuel and maintenance)	693,000 tons	0.16/ton	110,900	2.2
Electricity	3,636 kkW hr	0.029/kWh	105,444	2.1
Analyses	1,000 hr	17.00/man-hr	17,000	0.3
Subtotal conversion costs			1,791,500	34.9
Subtotal direct costs			2,650,900	51.7
Indirect costs				
Capital charges				
Depreciation, interim replacement, and insurance at 7.83% of total capital investment less land and working capital			672,500	13.1
Average cost of capital and taxes at 8.6% of total capital investment			831,500	16.2
Overhead				
Plant, 50% of conversion costs less utilities			843,000	16.4
Administrative, 10% of operating labor			133,150	2.6
Subtotal indirect costs			2,480,200	48.3
Total annual revenue requirements			5,131,100	100.0

	$/dry ton	$/wet ton	Mills/kWh
Equivalent unit revenue requirements	12.17	7.40	1.47

---

[a] Basis

Case I IUCS process with modified dewatering system. The thickener in the conventional dewatering train has been replaced by the proposed dewatering system (Auburn study).
Remaining plant life, 30 yr.
Coal burned, 482,571 lb/hr, 9,000 Btu/kWh.
Power plant on-stream time, 7,000 hr/yr.
Total capital investment, $10,717,000.
Midwest plant location, mid-1980 operating costs.

TABLE VI

Proposed System (Modified Dewatering; Case II)[a]

Total Annual Revenue Requirements - Regulated Utility Economics

	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Direct costs				
Delivered raw materials				
Lime (ground)	16,216 tons	53.00/ton	859,400	17.3
Subtotal raw materials cost			859,400	17.3
Conversion costs				
Operating labor and supervision				
Plant	26,280 man-hr	12.50/man-hr	328,500	6.6
Solids disposal equipment	52,560 man-hr	17.00/man-hr	893,500	18.0
Maintenance--plant labor and material, 4% of direct investment			179,800	3.6
Landfill operation				
Land preparation			11,000	0.2
Trucks (fuel and maintenance)	693,000 tons	0.06/ton	41,600	0.8
Earthmoving equipment (fuel and maintenance)	693,000 tons	0.16/ton	110,900	2.2
Electricity	3,506 kkW hr	0.029/kWh	101,674	2.0
Analyses	1,000 hr	17.00/man-hr	17,000	0.3
Subtotal conversion costs			1,684,000	33.9
Subtotal direct costs			2,543,400	51.2
Indirect costs				
Capital charges				
Depreciation, interim replacement, and insurance at 7.83% of total capital investment less land and working capital			677,600	13.6
Average cost of capital and taxes at 8.6% of total capital investment			835,400	16.8
Overhead				
Plant, 50% of conversion costs less utilities			791,200	15.9
Administrative, 10% of operating labor			122,200	2.5
Subtotal indirect costs			2,426,300	48.8
Total annual revenue requirements			4,969,700	100.0

	$/dry ton	$/wet ton	Mills/kWh
Equivalent unit revenue requirements	11.79	7.16	1.42

---

[a] Basis

Case II IUCS process with modified dewatering system. The thickener and filter in the conventional dewatering train has been replaced by the proposed dewatering system (Auburn study).

Remaining plant life, 30 yr.

Coal burned, 482,571 lb/hr, 9,000 Btu/kWh.

Power plant on-stream time, 7,000 hr/yr.

Total capital investment, $10,717,000.

Midwest plant location, mid-1980 operating costs.

TABLE VII

Base Case System (Conventional Dewatering)

Process Equipment List and Cost for Dewatering Area

Item	No.	Description	Equipment and material cost
1. Thickener and rake	1	219 ft diameter tank, 60 hp rake stainless steel or neoprene lined concrete basin with rake and motor (1 sp)	1,443,100
2. Pump, thickener overflow	2	881 gpm, 75 ft head, neoprene lined, 40 hp	11,400
3. Pump, thickener underflow	2	517 gpm, 75 ft head, neoprene lined, 25 hp	10,200
4. Rotary drum filter	3	464 ft^2, 10 ft diameter x 16 ft long drum stainless steel, vacuum and filtrate pumps	341,800
5. Pump, filtrate recycle	2	276 gpm, 75 ft head, neoprene lined, 10 hp motor	6,600
Total			1,813,100

TABLE VIII

Proposed System (Modified Dewatering Case I)

Process Equipment List and Cost for Dewatering Area

	Item	No.	Description	Equipment and material cost
1.	Thickener and rake	1	124 ft diameter tank, 20 hp rake stainless steel or neoprene lined concrete basin with rake and motor (1 sp)	761,800
2.	Pump, Lamella overflow	2	881 gpm, 75 ft head, neoprene lined, 40 hp	11,400
3.	Pump, thickener underflow	2	517 gpm, 75 ft head, neoprene lined, 25 hp	10,200
4.	Lamella clarifier	2	33,000 ft^2 area total plus tank	359,700
5.	Pump, Lamella cycle	2	494 gpm, 75 ft head, neoprene lined, 30 hp	10,200
6.	Rotary drum filter	3	464 ft^2, 10 ft diameter x 16 ft long drum stainless steel, vacuum and filtrate pumps	341,800
7.	Pump, filtrate recycle	2	276 gpm, 75 ft head, neoprene lined, 10 hp motor	6,600
	Total			1,501,700

TABLE IX

Proposed System (Modified Dewatering Case II)

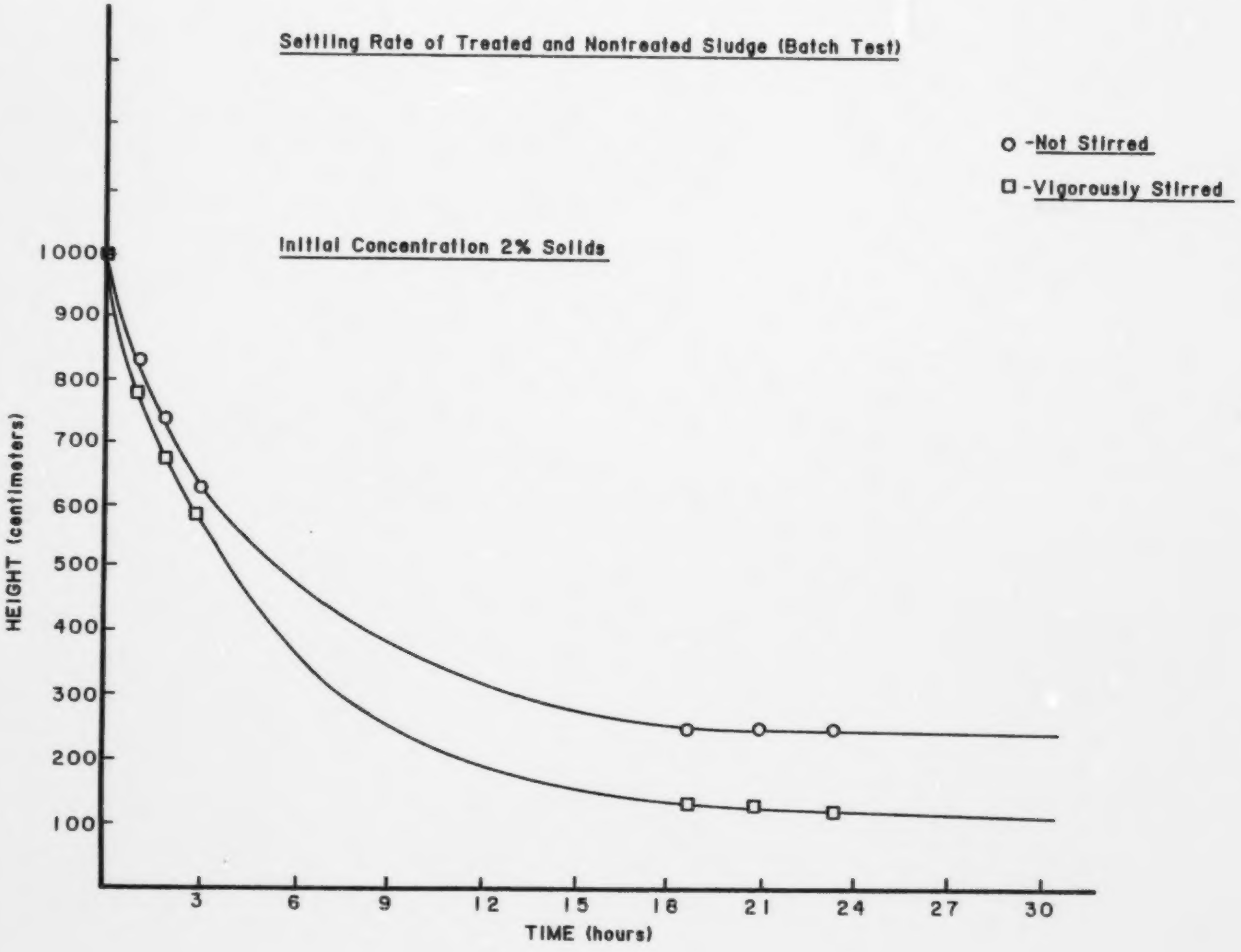
Process Equipment List and Cost for Dewatering Area

Item	No.	Description	Equipment and material cost
1. Thickener and rake	1	190 ft diameter tank, 20 hp, rake stainless steel or neoprene lined concrete with rake and motor (1 sp.)	1,199,300
2. Pump, Lamella overflow	2	972 gpm, 75 ft head, neoprene lined, 40 hp	11,400
3. Pump, thickener underflow	2	241 gpm, 75 ft, neoprene lined, 25 hp	10,200
4. Lamella clarifier	2	33,000 ft^2 area total plus tank	359,700
5. Pump, Lamella recycle	2	648 gpm, 75 ft head, neoprene lined, 30 hp	10,200
Total			1,590,800

Appendix D

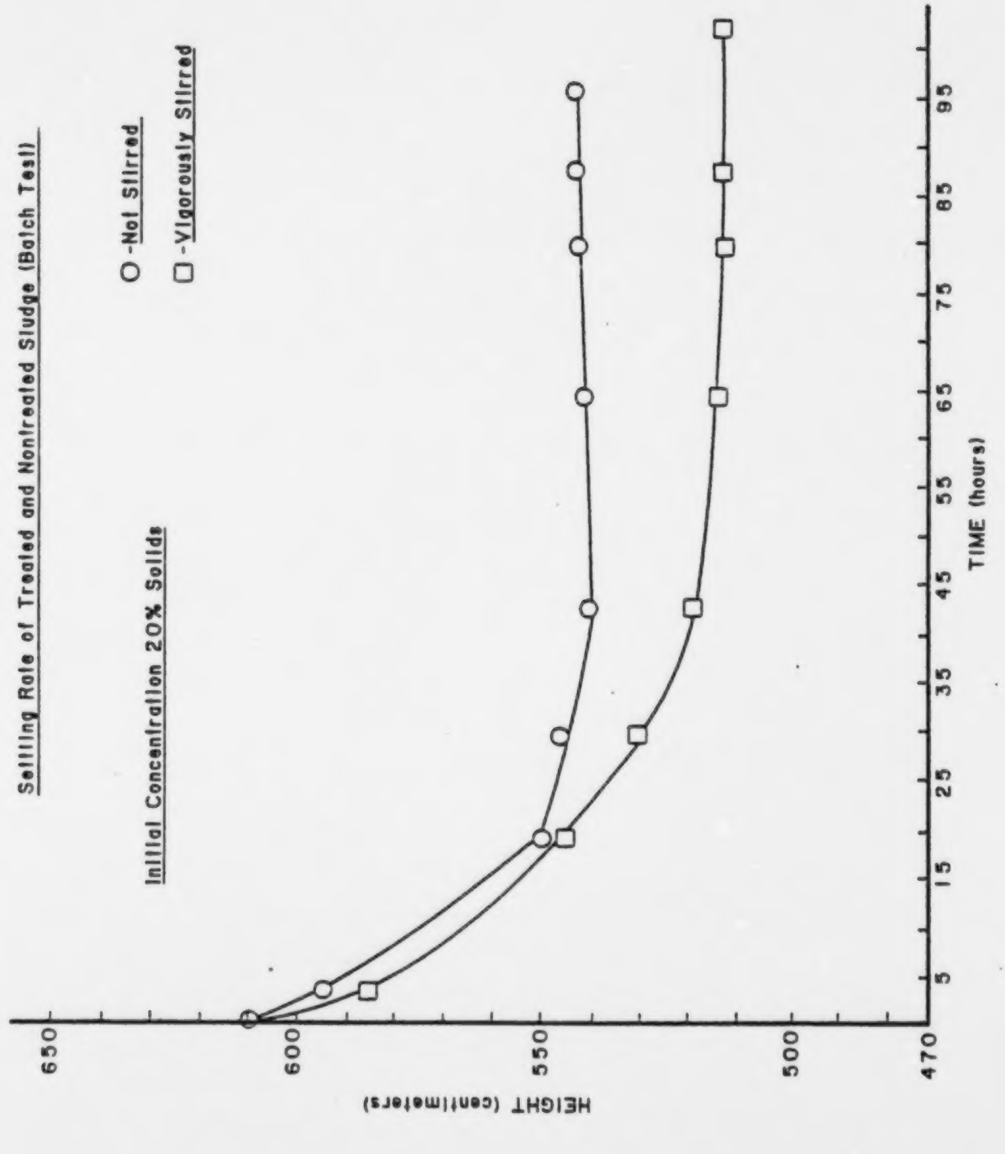
Effects of Stirring on Settling Behavior of FGD Wastes

1. Batch Tests
2. Continuous Tests

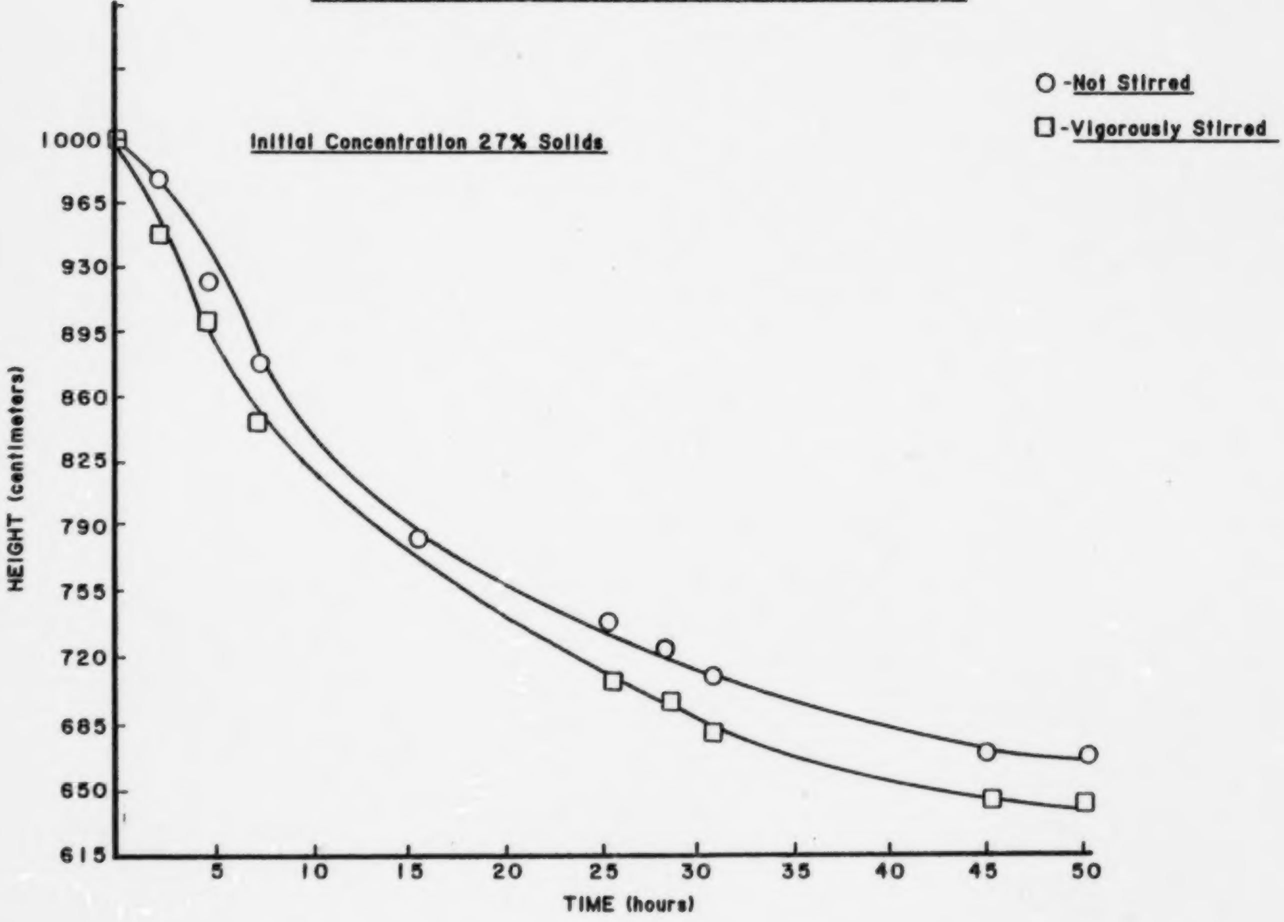


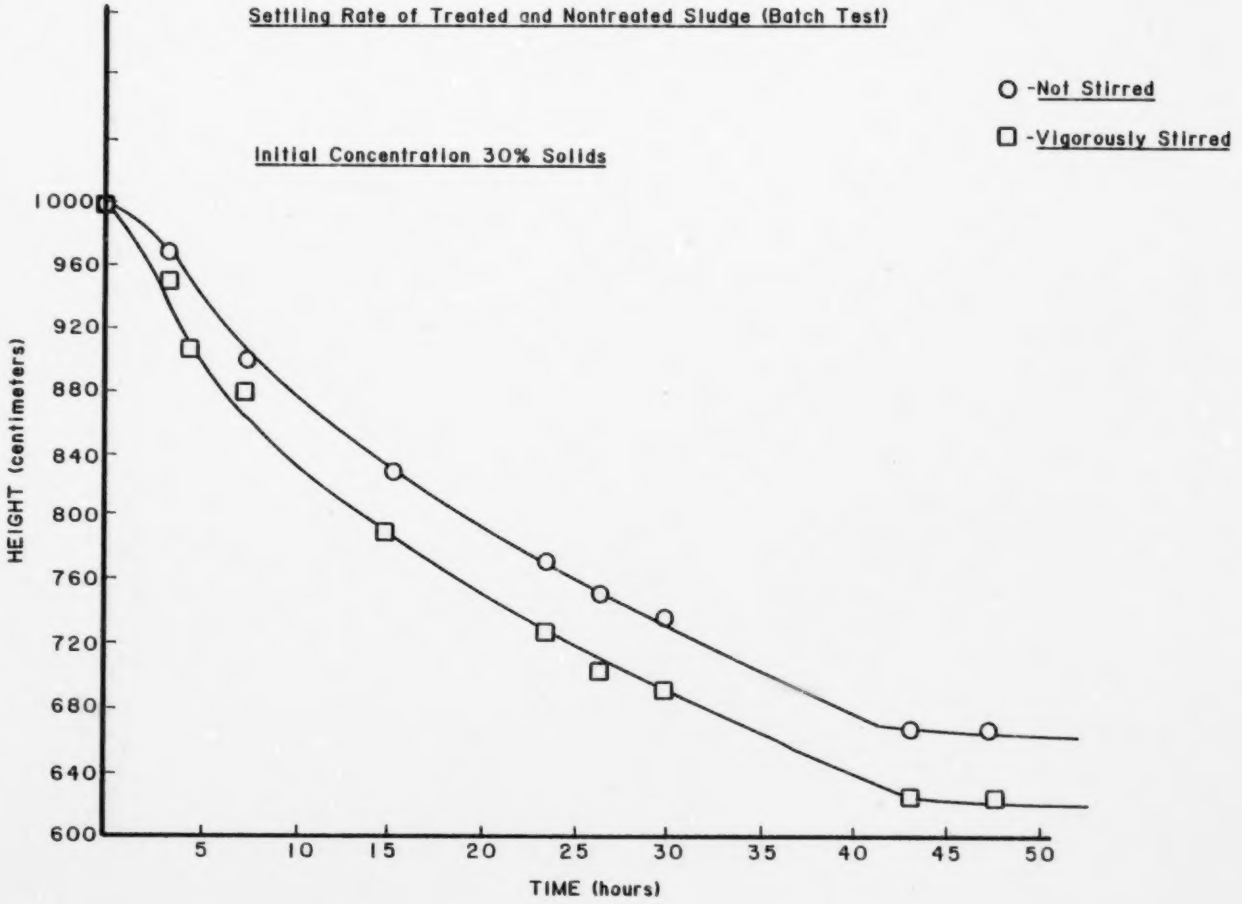
Settling Rate of Treated and Nontreated Sludge (Batch Test)

Initial Concentration 2% Solids


Settling Rate of Treated and Nontreated Sludge (Batch Test)
Initial Concentration 20% Solids
○ -Not Stirred
□ -Vigorously Stirred
HEIGHT (centimeters)
650
600
550
500
470
TIME (hours)
5
15
25
35
45
55
65
75
85
95

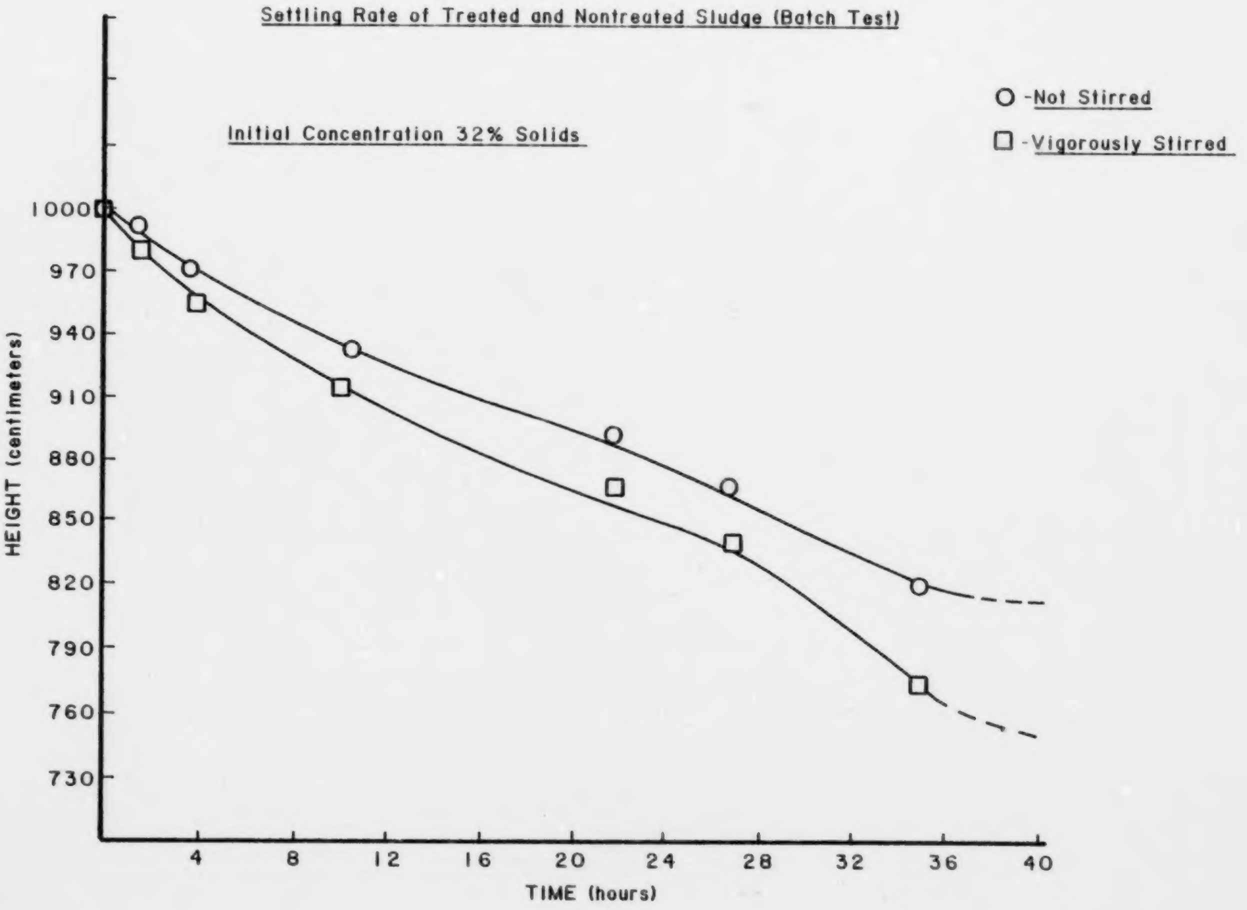
Settling Rate of Treated and Nontreated Sludge (Batch Test)



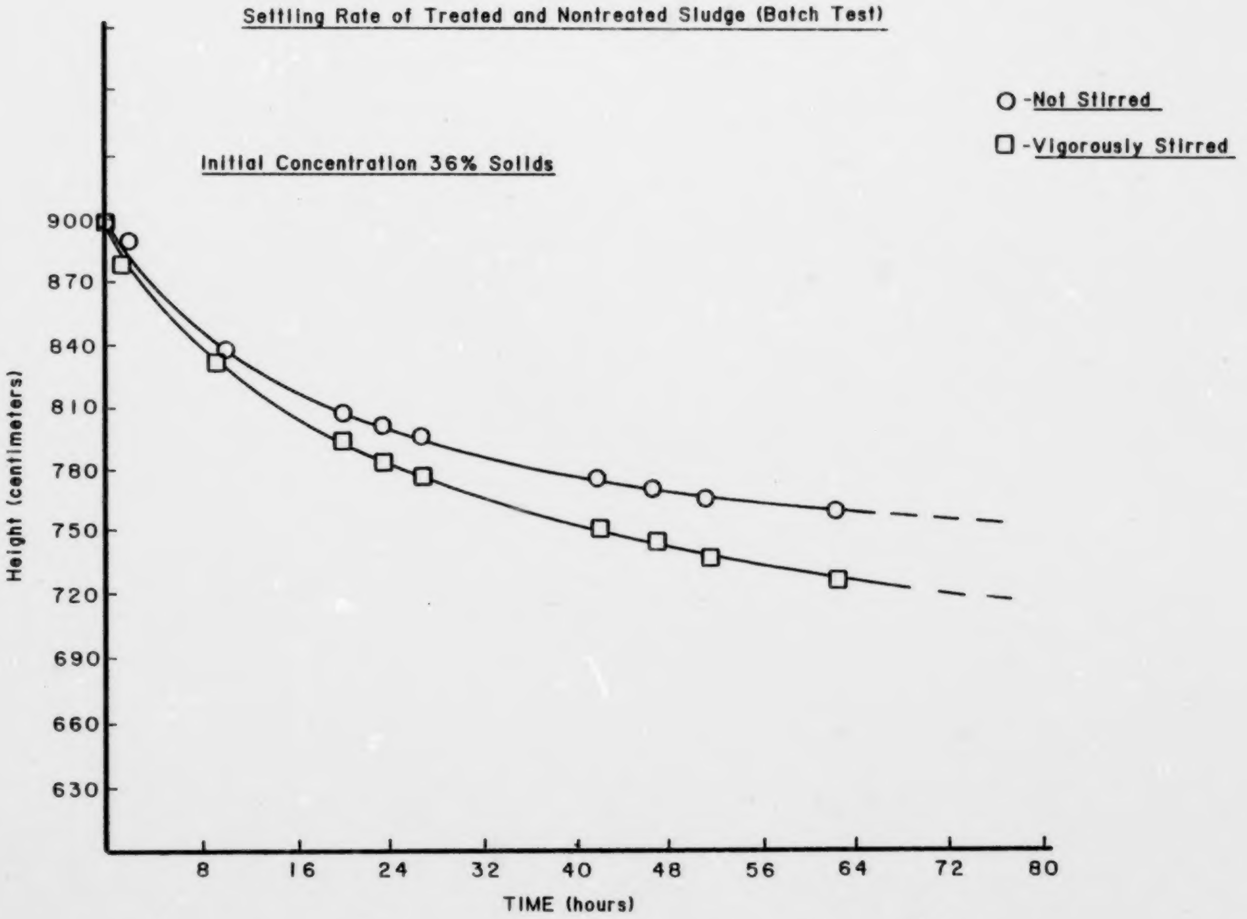


Settling Rate of Treated and Nontreated Sludge (Batch Test)

Initial Concentration 30% Solids

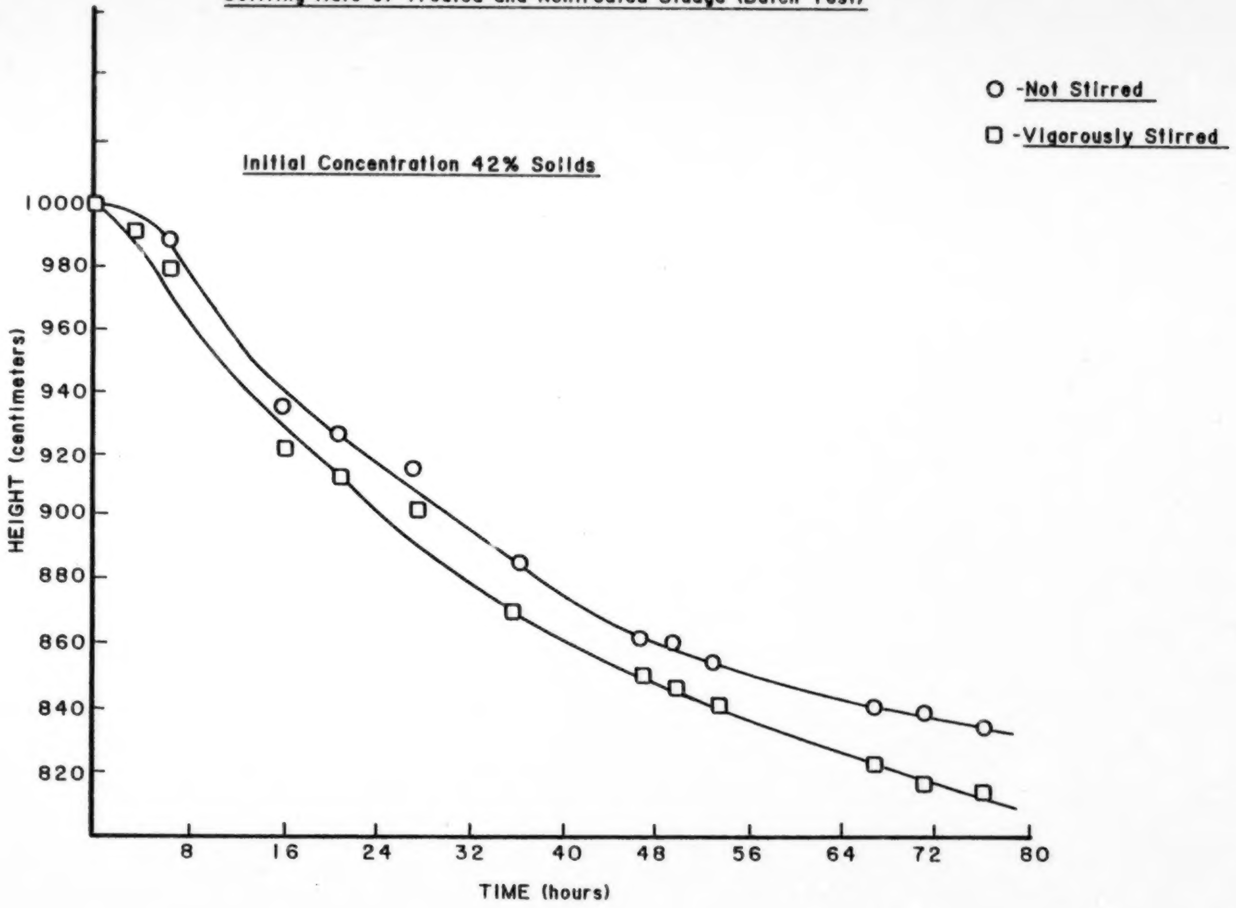


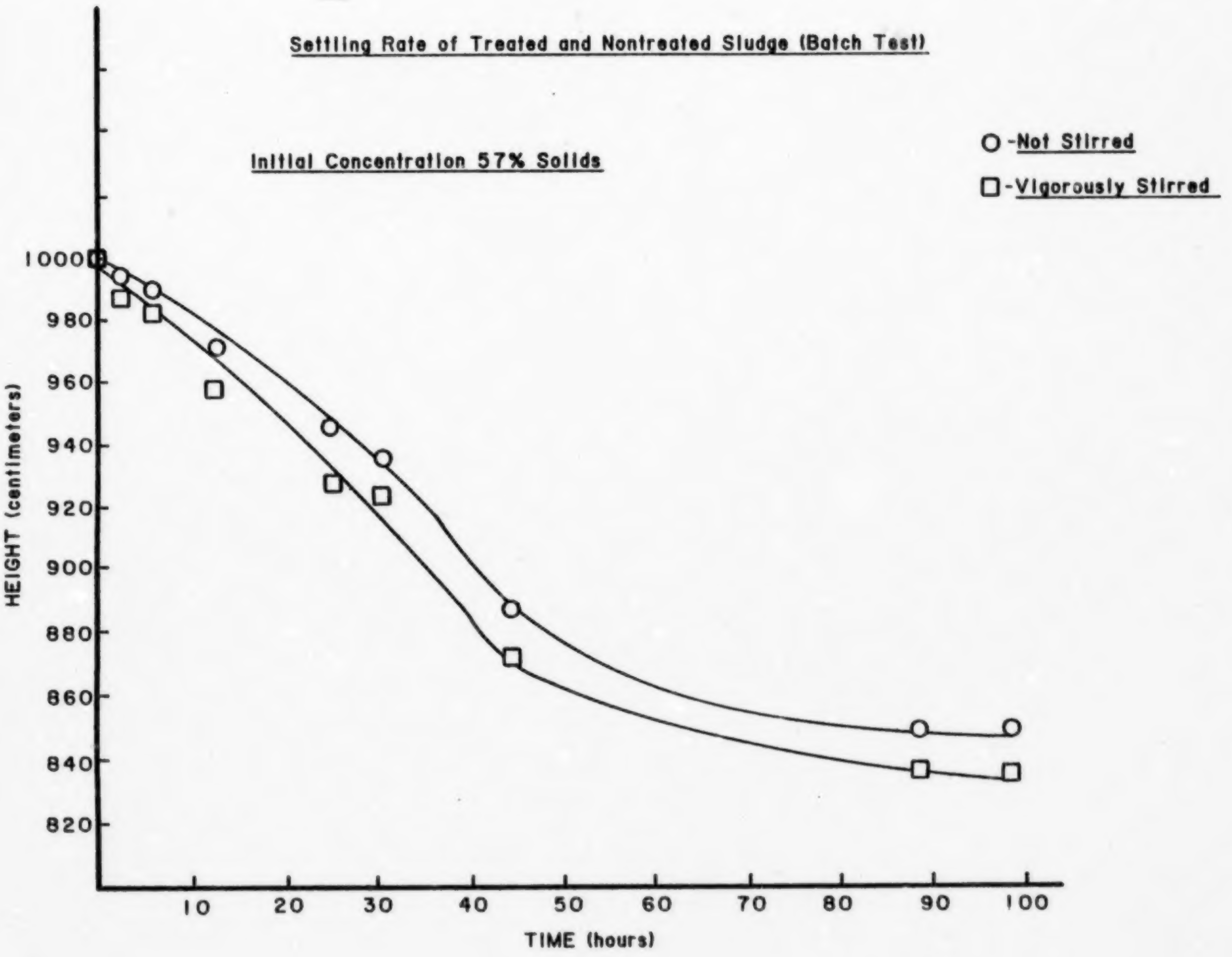
Settling Rate of Treated and Nontreated Sludge (Batch Test)



Settling Rate of Treated and Nontreated Sludge (Batch Test)

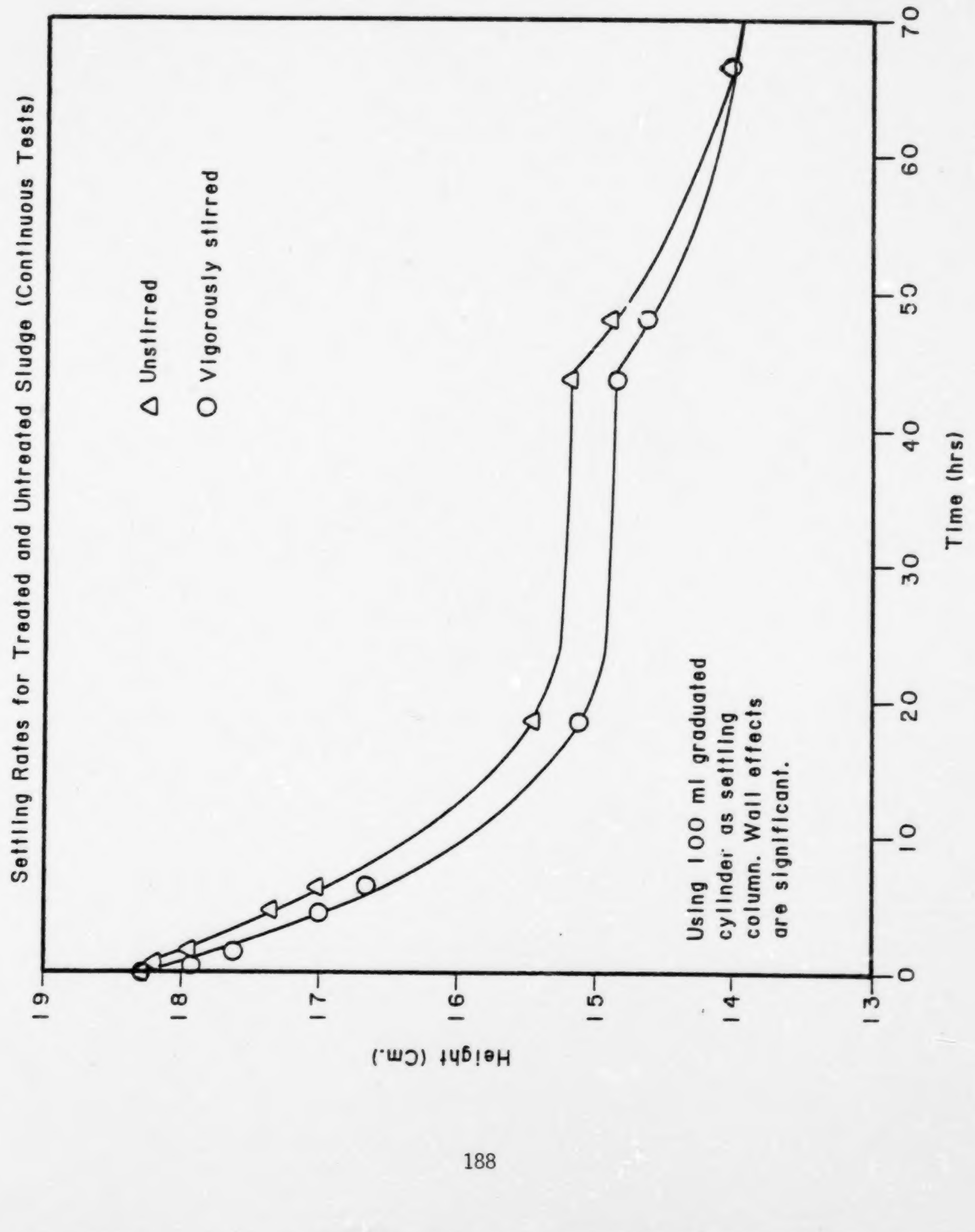
Settling Rate of Treated and Nontreated Sludge (Batch Test)



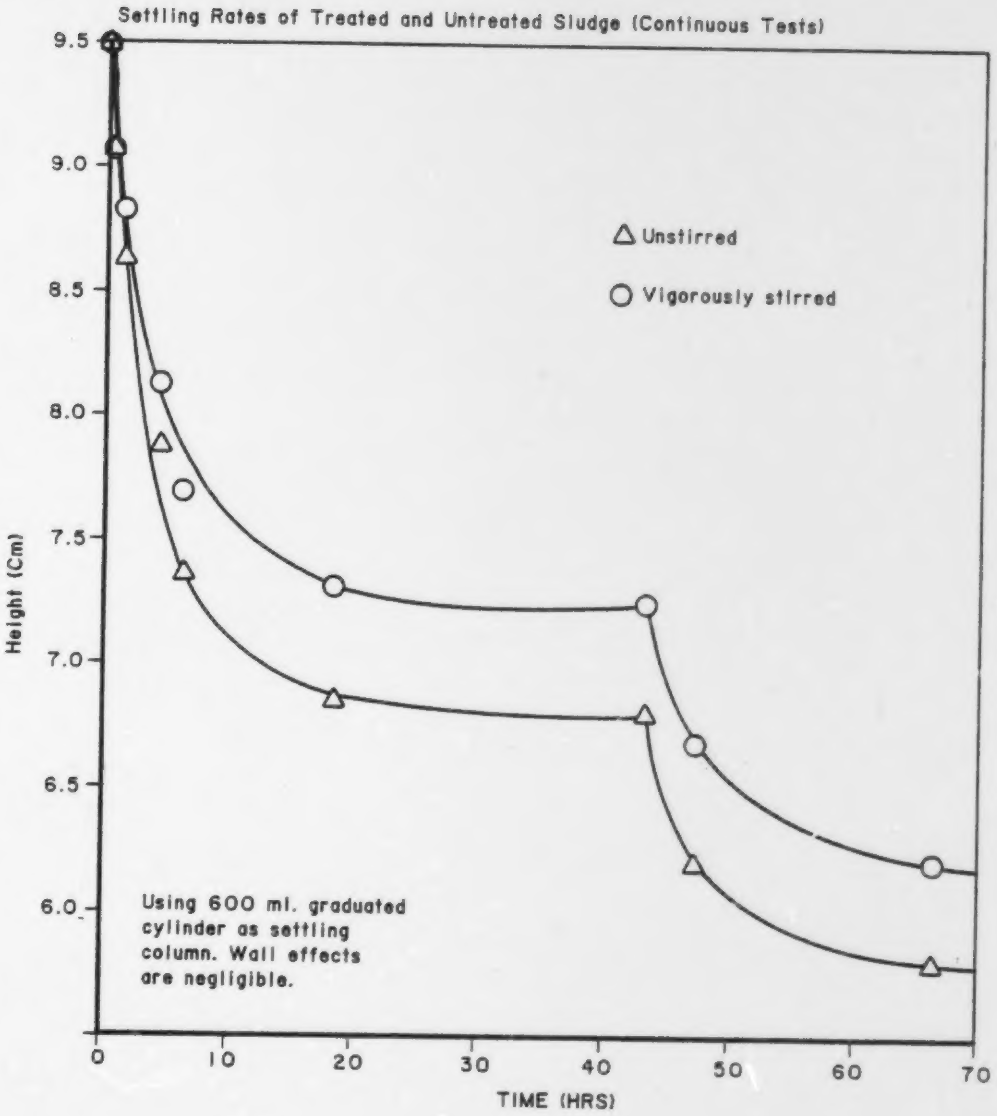


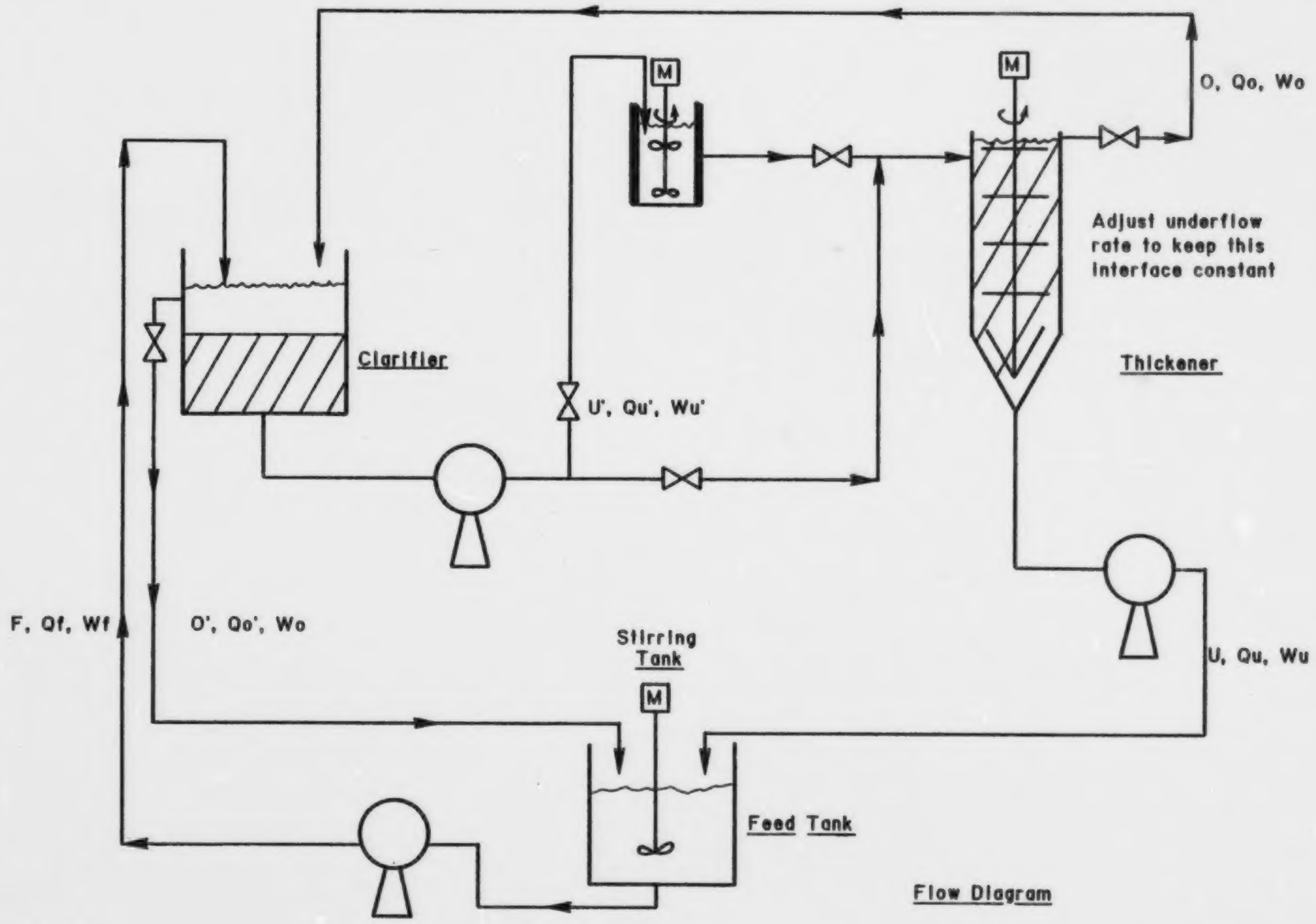
Settling Rate of Treated and Nontreated Sludge (Batch Test)

Initial Concentration 57% Solids



Settling Rates for Treated and Untreated Sludge (Continuous Tests)


Settling Rates of Treated and Untreated Sludge (Continuous Tests)
△ Unstirred
○ Vigorously stirred
Height (Cm)
9.5
9.0
8.5
8.0
7.5
7.0
6.5
6.0
0
10
20
30
40
50
60
70
TIME (HRS)
Using 600 ml. graduated cylinder as settling column. Wall effects are negligible.


O, Qo, Wo
Adjust underflow rate to keep this interface constant
Thickener
Clarifier
U', Qu', Wu'
F, Qf, Wf
O', Qo', Wo
Stirring Tank
U, Qu, Wu
Feed Tank
M
M
M

Flow Diagram

Effects of Stirring on Settling Behavior

Results

I. Control Sample (Unstirred)

Flow rate of feed, Q_F = 150 ml/min

Flow rate of underflow from the thickener, Q_u = 44 ml/min

Flow rate of overflow from the thickener, Q_o = 3.0 ml/ml

Flow rate of underflow from the clarifier, Q'_u = 47.0 ml/min

Flow rate of overflow from the clarifier, Q'_o = 106 ml/min

Concentration	Date: 12/21	12/22	12/23	12/26	12/27	Avg.
W_u	35.95	36.43	32.73	31.10	31.16	31.77
W_F	8.64	8.50	9.36	8.28	8.98	9.325

$$\Delta W = W_u - W_F = 31.77 - 9.325 = 22.45 \text{ wt\%}$$

II. Experimental Sample (Stirred)

Flow rate of feed, Q_F = 155.0 ml/min

Flow rate of underflow from the thickener, Q_u = 37.25 ml/min

Flow rate of overflow from the thickener, Q_o = 7.25 ml/min

Flow rate of underflow from the clarifier, Q'_u = 44.5 ml/min

Flow rate of overflow from the clarifier, Q'_o = 117.25 ml/min

Concentration	Date: 1/19	1/21	1/22	1/23	1/24	Avg.
W_u	38.51	30.82	30.63	29.21	31.08	30.435
W_F	6.36	7.60	6.17	8.23	8.12	7.53
W_u^1	34.90	27.17	25.16	---	26.05	26.13

$$\Delta W = W_u - W_F = 30.435 - 7.53 = 22.905 \text{ wt\%}$$

Appendix E

Details of Second Economic Evaluation by Engineering Staff of Tennessee Valley Authority

Test Plan for Operation of
Auburn Dewatering Unit

A. START - UP

1. Prefill 30% of the thickener with material from bottom of existing clarifier.
2. Finish prefilling the thickener with slurry feed at a fast rate.
3. Once the thickener overflows, cut the slurry feed rate back to 4 gpm.
4. Once clarifier overflows, start recirculation at a rate approximately the same as the feed rate.
5. Start the thickener underflow pump at the setting given in Figure 1 (0.82 gpm).

B. RUN 1

1. Objective: Obtain highest underflow concentration (product quality) while maintaining an acceptable overflow clarity for a set inlet feed rate (15% / 4 gpm).
2. Initial Settings
    a. The underflow rate corresponding to a desired underflow concentration can be calculated from an overall solids balance as below:

$$Q_u = \frac{Q_f C_f - Q_o C_o}{C_u},$$

where

$$C = \frac{W/\rho_s}{W/\rho_s + (1-W)/\rho_1}$$

    b. The clarifier underflow rate can be ratioed to the feed slurry flow rate.

    $Q_{uo} - R_2 Q_f$

    Initially R_2 will be taken as unity.

3. Operation Plan
    a. Action plans
        1. Unclear overflow
            a. Adjust the recycle (clarifier underflow) pump as a primary control to maintain overflow clarity and to maintain maximum clarifier underflow concentration.

   b. Adjust the underflow pump as a secondary control to maintain overflow clarity and to maintain maximum underflow concentration.
2. Underflow to lean: decrease the underflow pump rate.
3. Underflow pump failure
   a. Shut off feed
   b. Valve off the underflow pump.
   c. Romove the underflow pump for repair.
   d. Dump the thickener contents.
4. Recycle pump failure
   a. Shut off feed.
   b. Valve off the clarifier.
   c. Remove the recycle pump for repair.
   d. Recycle the thickener.
5. Failure of sump tank
   a. Shut off feed
   b. Recycle thickener
   c. Repair sump tank

b. Solids mass balance (See Table 1 Analytical Schedule)
1. Flowrates and percent total solids
   a. The flow rate can be measured at each point in the system.
   b. Percent total solids can be determined by drying on a moisture determination balance for 15 to 20 minutes.
2. Approach to steady state
   a. In initial approach to steady state only measure the thickener underflow (stream 4) concentration (minimum six times per day.)
   b. Overall solids balance: Minimum twice a day; preferably four times a day. Measure the clarifier overflow (stream 3), underflow (stream 4) and slurry feed (stream 1) flow rate and solids content. The deviation from steady state (DSS) is then calculated as follows:

$$DSS = \left(1 - \frac{Q_u C_u + Q_o C_o}{Q_f C_f} \right) 100\%$$

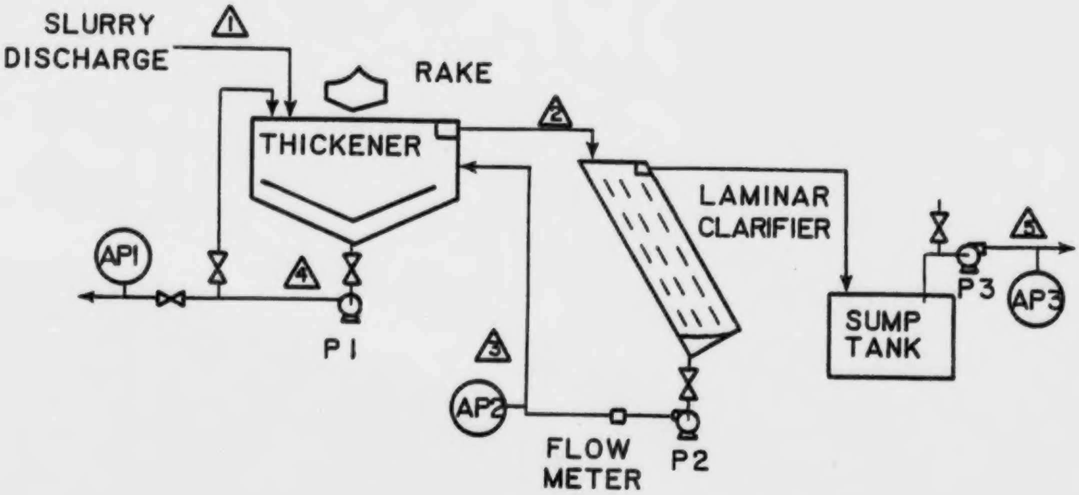
When DSS is reasonably small steady state has been reached.

   c. Unit solids balances
      1. All the concentrations in the system should be measured 4 times six hours apart to make certain steady state has been reached.
      2. Efficiency and settling flux for overall system and each unit can be calculated from measurements of concentration and flow rate of all the streams.

c. Feed characterization: Determine the feed concentration and the settling rate (minimum twice per day; preferably four times per day.)

FIGURE 1

SCHEMATIC OF AUBURN DEWATERING UNIT


SLURRY DISCHARGE
1
RAKE
THICKENER
2
LAMINAR CLARIFIER
5
AP1
4
P1
3
AP2
FLOW METER
P2
SUMP TANK
P3
AP3

STREAM NO.	1	2	3	4	5
DESCRIPTION	Slurry Discharge	Thickener Overflow	Clarifier Underflow	Thickener Underflow	Clarifier Overflow
Rate, lb/hr.	2186	3923	2332	595	1591
RATE, gpm	4.00	7.18	4.00	0.82	2.18
% SOLIDS	15.0	14.9	25.0	55.0	——
SPEC. GRAVITY	1.093	1.092	1.165	1.451	1.000

Based on $\rho_s = 2.3$ gm ml^{-1}, $\rho_l = 1.0$ gm ml^{-1}

TABLE 1

ANALYTICAL SCHEDULE

Number of times analysis performed per day.

Purpose	Stream Nos.	Percent Total Solids	Flow Rate	Settling Rate
Feed Characterization	1	4	—	4
Initial Approach to Steady State	4	6	—	—
Overall Mass Balance	1,4,5	4	4	—
Unit Solids* Balances	1,2,3,4,5	4	4	—

*After steady state has been reached.

C. RUN 2 and SUBSEQUENT RUNS

1. Objective: Check to see if system was operating near the maximum throughput during the previous run.
2. Initial settings
    - a. Set feed flow rate for 2 gpm greater than previous run.
    - b. Underflow pump setting: set to yield final percent solids obtained in previous run.
    - c. Recycle pump: ratio of recycle to feed flow rate set equal to the final value used in the previous run.
3. Operation plan: same as run 1.

NOMENCLATURE

C	Volume fraction of solids.
DSS	Deviation from steady state, percent.
Q_f	Flow rate of feed to the thickener, gpm.
Q_{OT}	Flow rate of thickener overflow, gpm.
Q_{UC}	Flow rate of clarifier underflow, gpm.
Q_U	Flow rate of thickener underflow, gpm.
Q_O	Flow rate of clarifier overflow, gpm.
R_1	Q_u/Q_f.
R_2	Q_{UC}/Q_f.
W	Weight fraction of solids.
ρ_1	Density of clear liquid, gm ml^{-1}.
ρ_s	Density of solids, gm ml^{-1}.

Appendix F

Details of Second Economic Evaluation by Engineering Staff of Tennessee Valley Authority

J. L. Crowe, Projects Manager, Waste Water and Solid Waste Control,
Environmental Control and Assessment Branch, Energy Technology, Division
of ED&T, 1140 CST2-C
H. L. Faucett, Projects Manager, Muscle Shoals Operations Office, ED&O,
Division of ED&T, 501 CEB-M
March 4, 1982

AUBURN SLUDGE DEWATERING EVALUATION

As requested during our meeting on December 17, the Muscle Shoals Operations Office has prepared an economic evaluation of the Auburn dewatering system for FGD sludge. Also, an evaluation of the physical condition of the Auburn dewatering system pilot plant located at the Shawnee Test Facility and the comparative economics of testing this equipment at Shawnee versus Widows Creek were completed. Cost estimates for these evaluations were prepared by the Pilot-Plant Design and Construction Branch.

With the assistance of Steve Newton, Shawnee Test Facility, it was estimated that the Auburn dewatering system pilot-plant equipment located at Shawnee can be made operable with an expenditure of about $6,700 (materials, labor, and engineering). All of the components of the system except for some slurry hose are onsite and can be made mechanically suitable for operations with maintenance and minor repairs.

Relocation of the system to the Widows Creek Steam Plant will require an expenditure of $63,900 (study-grade estimate) as itemized in the attached Table I (Attachment A). The equipment can be located adjacent to the existing forced-oxidation pilot-plant equipment in the unit 8 scrubber area. Feed to the system can be taken from the unit 8 scrubber train that is equipped for forced-oxidation testing. The small building containing electrical equipment for the forced-oxidation pilot plant can be used for housing process control and electrical equipment for the Auburn dewatering system. Product from the test equipment can be returned to the unit 8 scrubber effluent hold tank.

The economic evaluation of a full-scale (1125 MW) Auburn dewatering system (Attachment B) was performed by comparing it with a conventional thickener/filter system. System boundaries for the evaluation were defined to permit only the comparison of the dewatering system areas of a limestone scrubbing process (excluding landfill). Three cases were evaluated: (1) conventional dewatering system-base case with a thickener and filter, (2) Auburn dewatering system - Case I with a thickener-clarifier combination, and (3) Auburn dewatering system - Case II with a thickener-clarifier combination followed by a filter. The dewatering systems were designed to increase the concentration of the oxidized slurry discharge from Widows Creek units 7 and 8 scrubbers from 12% to 80% solids by weight.

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Attachment A

TABLE I

Cost Estimate

Relocation of Auburn Dewatering Equipment to Widows Creek Steam Plant

Pilot-Plant Design and Construction Branch Activities:	
Design and detail engineering	$ 4,000
Drafting	3,000
Procurement	800
Cost estimating	500
Construction supervision	3,500
Travel and per diem	1,600
Subtotal	13,400
Accuracy factor (+40%)	5,400
Total	$18,800
Construction Activities:	
Process equipment	$ 1,800
Piping and valves	1,200
Electrical	1,200
Instrumentation	1,000
Site preparation and foundations	2,300
Equipment rental	3,700
Miscellaneous	2,000
Construction labor	19,000
Subtotal	32,200
Accuracy factor (+40%)	12,900
Total	$45,100
Summary:	
Pilot-Plant Design and Construction Branch Activities	$18,800
Construction Activities	45,100
Total	$63,900

AUBURN SLUDGE DEWATERING EVALUATION

The total capital investments and annual revenue requirements are summarized as follows:

	Total capital investment		Annual revenue requirement
	k$	$/kW	k$
Conventional-base case	6,076	5.40	1,408
Auburn - Case I	2,486	2.21	710
Auburn - Case II	4,801	4.27	1,164

The capital savings for using an Auburn dewatering system (Case I) instead of the conventional dewatering system is about $3.6M or $3.2/kW. This amount is equivalent to about 1.5% of the total capital investment for a limestone scrubber process. It should be noted that the specified premise in Case I of a thickener underflow concentration of 80% solids by weight remains to be demonstrated. However, the Auburn dewatering system Case II) may be used to dewater the scrubber slurry discharge to a concentration of 65% solids by weight before filtering the slurry to produce a gypsum filter cake with a concentration of 80% solids by weight. Capital savings of about $1.2M or $1.1/kW could still be achieved. Annual revenue savings of $240,000 to $700,000 are projected for use of the Auburn dewatering system versus the conventional dewatering system.

If you have any questions or comments, please call Jack Henson at extension 2938.

LJH:KF
Attachments
cc (Attachments):
- ARMS, 640 CST2-C
- J. W. Barrier, 501 CEB-M
- H. W. Elder - G. G. McGlamery, 501 CEB-M
- H. B. Flora II, 1140 CST2-C
- L. J. Henson, 501 CEB-M
- G. A. Hollinden, 1130 CST2-C
- S. B. Jackson, 501 CEB-M
- J. W. Jones, 1140 CST2-C
- W. E. O'Brien, 501 CEB-M

Attachment B

ECONOMIC EVALUATION OF AUBURN DEWATERING SYSTEM

An economic evaluation of the Auburn dewatering system was performed by comparing it with a conventional thickener/filter system. System boundaries for the evaluation were defined to permit only the comparison of the dewatering system areas of a limestone scrubbing process (excluding landfill). The Auburn and conventional systems were assumed to dewater a slurry discharge stream from the Widows Creek coal-fired steam plant (units 7 and 8) scrubbers. A gypsum waste product suitable for landfill was produced by increasing the concentration in the slurry discharge from 12% to 80% solids by weight. Flow diagrams and material balances for the three cases considered in the evaluation are shown in Figures 1, 2, and 3. The three cases are described as follows:

1. Conventional Dewatering System - Base Case: The oxidized slurry discharge from the scrubber system is thickened in a conventional thickener to a concentration of 40% solids by weight and pumped to rotary drum filters. The filter cake is discharged from the filters at a concentration of 80% solids by weight. Thickener overflow liquor and filtrate are returned to the scrubber system.

2. Auburn Dewatering System - Case I: The oxidized slurry discharge is pumped to a thickener which overflows to a Lamella-type clarifier. The slurry underflow from the clarifier with a concentration of 40% solids by weight is pumped back to the thickener and injected below the surface. The concentration of solids in the slurry underflow from the thickener is 80% by weight. The clarifier overflow liquor is returned to the scrubber system.

3. Auburn Dewatering System - Case II: This case is similar to Case I except that the thickener is designed to yield an underflow concentration of 65% solids by weight. The thickener underflow slurry is pumped to rotary drum filters which provide a filter cake for sludge disposal with a concentration of 80% solids by weight. The clarifier overflow liquor and filtrate are returned to the scrubber system.

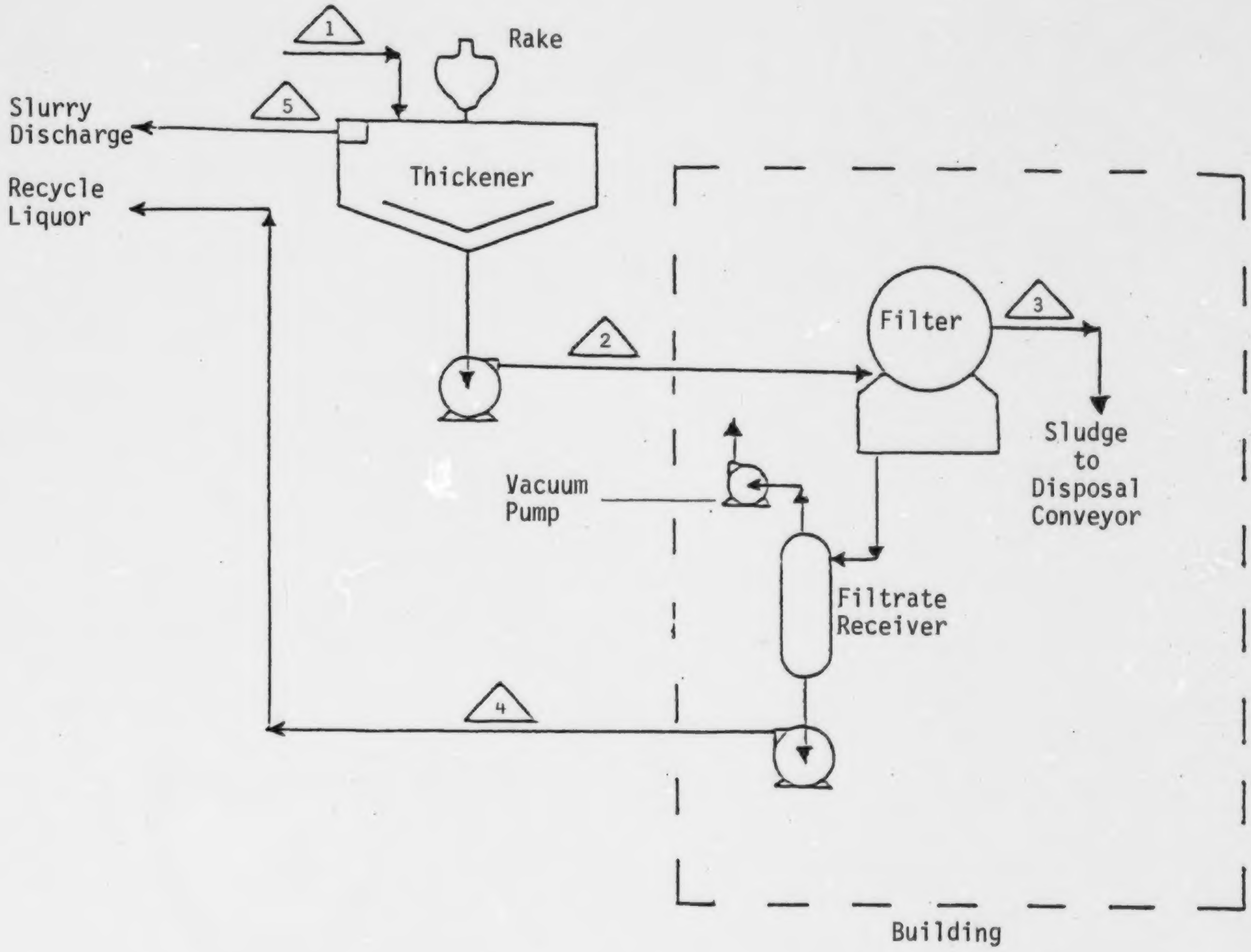
The material balances and equipment specifications were based on the design premises shown in Table I for Widows Creek units 7 and 8 scrubbers. The unit area requirements for the thickener and filter in the conventional dewatering system were based on previous test results from projects at the Widows Creek and Shawnee Steam Plants. These factors and additional settling rate data were given to Dr. Ray Tarrer of Auburn University for sizing the thickener and clarifier in the Auburn dewatering system. Test data were not available for filtering a slurry with a concentration of 65% solids by weight to produce a filter cake with a concentration of 80% solids by weight (Case II). However, the filters were sized by extrapolating previous Shawnee data and using basic engineering principles. Unit area requirements for dewatering equipment used in each case are presented in Table II.

Specifications and costs for major equipment items in each of the systems are presented in Tables III, IV, and V. Estimates of the equipment costs are based on TVA cost data and vendor quotations.

The estimated process equipment costs were used to calculate total capital investment and annual revenue requirements for each case. Cost breakdowns for each of the cases are presented in Tables VI through XI. The total capital investment and annual revenue requirements are summarized as follows:

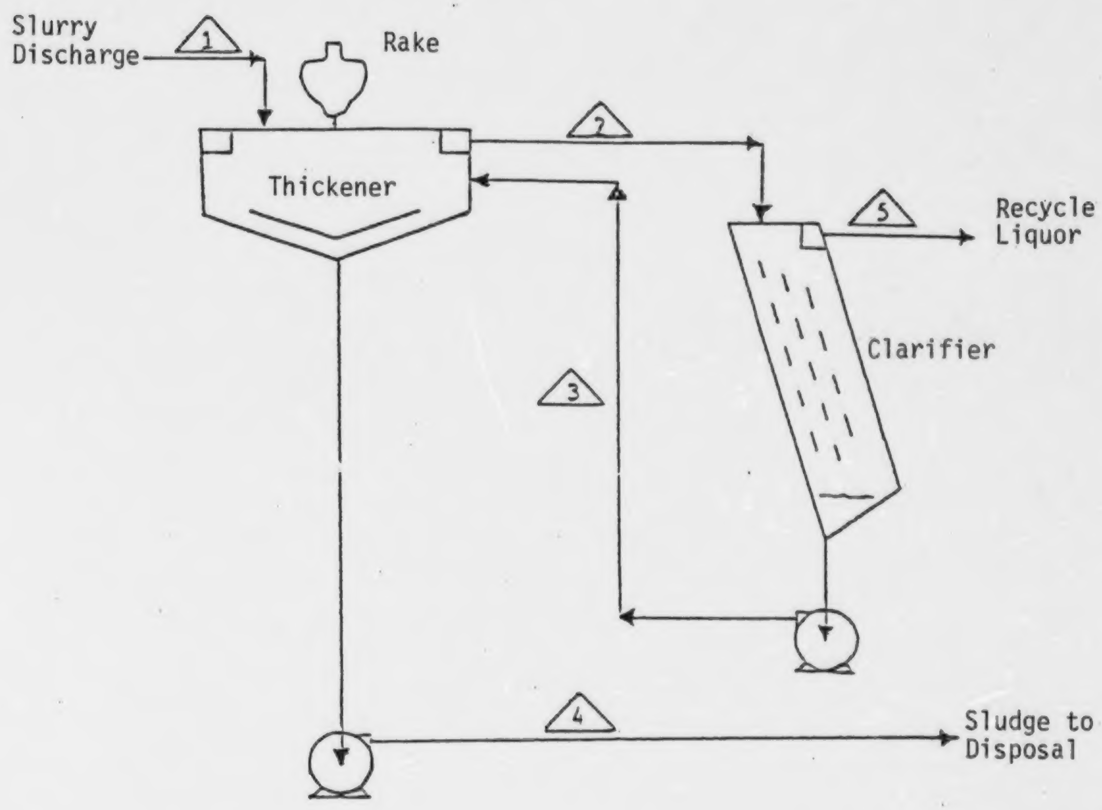
	Total capital investment		Annual revenue requirement
	k$	$/kW	k$
Conventional - base case	6,076	5.40	1,408
Auburn - Case I	2,486	2.21	710
Auburn - Case II	4,801	4.27	1,164

Figure 1. Flow Diagram and Material Balance. Conventional Dewatering System - Base Case.


1
Rake
Slurry Discharge
5
Thickener
Recycle Liquor
Filter
3
2
Sludge to Disposal Conveyor
Vacuum Pump
Filtrate Receiver
4
Building

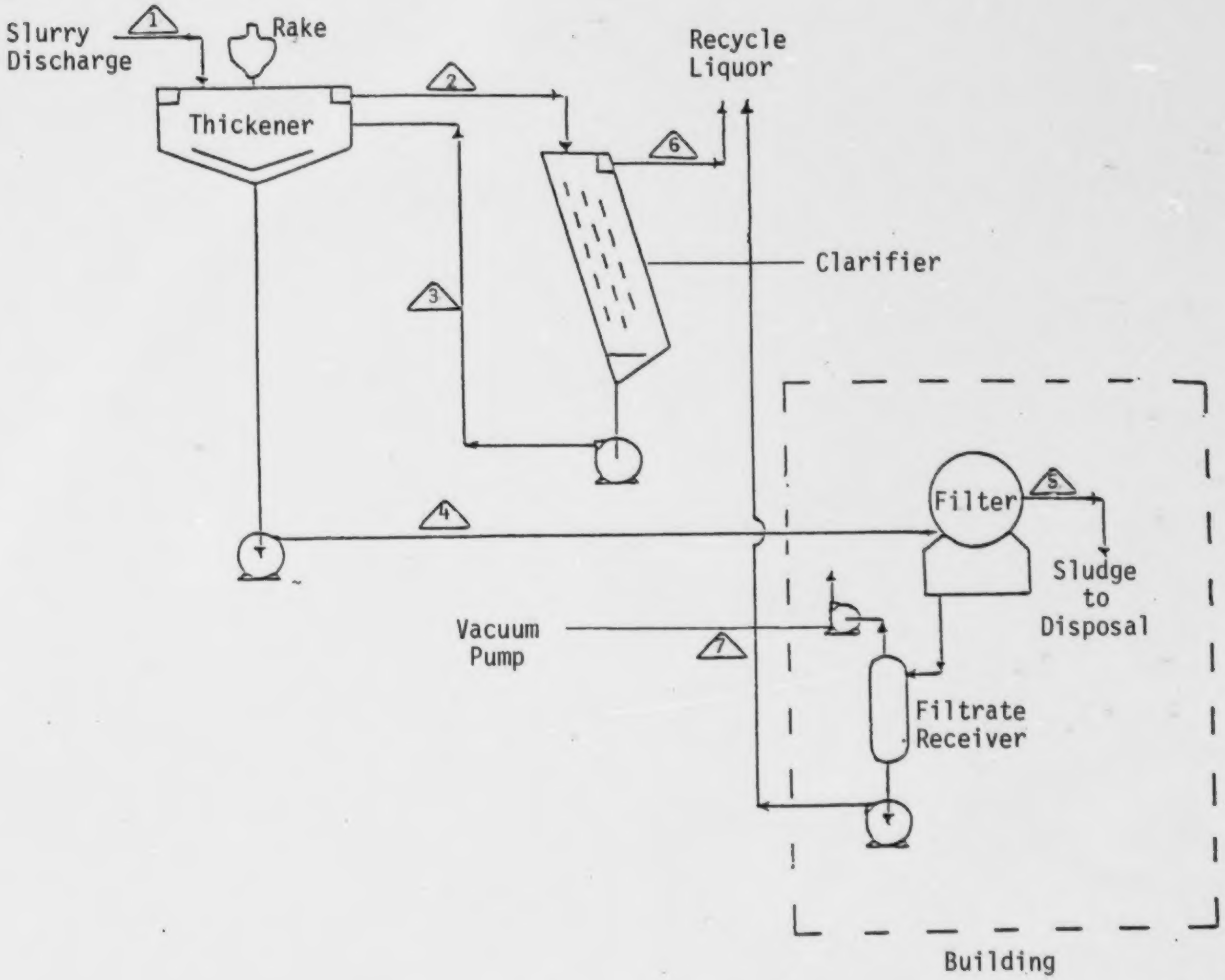
STREAM No.	1	2	3	4	5
DESCRIPTION	Slurry Discharge	Thickener Underflow	Sludge to Disposal	Filtrate	Thickener Overflow
RATE, lb./hr.	2,536,167	760,850	380,425	380,425	1,775,317
RATE, gpm	4,693	1,152	394	758	3,541
% SOLIDS	12	40	80	-	-
SPEC. GRAVITY	1.08	1.32	1.93	1.00	1.00

Figure 2. Flow Diagram and Material Balance. Auburn Dewatering System -Case I


Slurry Discharge
1
Rake
2
Thickener
5
Recycle Liquor
Clarifier
3
4
Sludge to Disposal

STREAM No.	1	2	3	4	5
DESCRIPTION	Slurry Discharge	Thickener Overflow	Clarifier Underflow	Sludge to Disposal	Clarifier Overflow
RATE, lb./hr.	2,536,167	2,916,592	760,850	380,425	2,155,742
RATE, gpm	4,693	5,451	1,152	394	4,299
% SOLIDS	12	-	40	80	-
SPEC. GRAVITY	1.08	-	1.32	1.93	1.00

Figure 3. Flow Diagram and Material Balance. Auburn Dewatering System - Case II.


Slurry Discharge
1
Rake
Thickener
2
Recycle Liquor
6
Clarifier
3
4
Filter
5
Sludge to Disposal
Vacuum Pump
7
Filtrate Receiver
Building

STREAM No.	1	2	3	4	5	6	7
DESCRIPTION	Slurry Discharge	Thickener Overflow	Clarifier Underflow	Thickener Overflow	Sludge to Disposal	Clarifier Overflow	Filtrate
RATE, lb./hr.	2,536,167	2,828,802	760,850	468,215	380,425	2,067,952	87,790
RATE, gpm	4,693	5,274	1,152	571	394	4,122	177
% SOLIDS	12	-	40	65	80	-	-
SPEC. GRAVITY	1.08	-	1.32	1.64	1.93	1.00	1.00

TABLE I

Design Premises

<table>
<tr><th></th><th>Unit 7</th><th>Unit 8</th></tr>
<tr><td>Widows Creek Steam Plant Parameters</td><td></td><td></td></tr>
<tr><td>Rated load, kW</td><td>575,000</td><td>550,000</td></tr>
<tr><td>Heat rate, Btu/kWh</td><td>9,360</td><td>9,500</td></tr>
<tr><td>Fly ash overhead, %</td><td>80</td><td>80</td></tr>
<tr><td>Fly ash collector efficiency, %</td><td>80</td><td>20</td></tr>
<tr><td>Sulfur overhead, %</td><td>95</td><td>95</td></tr>
<tr><td>Coal Characteristics</td><td></td><td></td></tr>
<tr><td>Higher heating value, Btu/lb</td><td colspan="2">11,500</td></tr>
<tr><td>Ash content, %</td><td colspan="2">9-15</td></tr>
<tr><td>Sulfur content, %</td><td colspan="2">2.8-4.5</td></tr>
<tr><td>Moisture content, %</td><td colspan="2">7.8-9.7</td></tr>
<tr><td>Scrubber Operating Parameters</td><td></td><td></td></tr>
<tr><td>Limestone stoichiometry, lb-mols $CaCO_3$/lb-mol (inlet) SO_2</td><td>1.25</td><td>1.50</td></tr>
<tr><td>Limestone inerts, %</td><td colspan="2">4.5</td></tr>
<tr><td>SO_2 removal efficiency, %</td><td colspan="2">85</td></tr>
<tr><td>Fly ash removal efficiency, %</td><td colspan="2">99</td></tr>
<tr><td>Oxidation, lb-mols sulfate/lb-mol SO_2 removed</td><td colspan="2">95</td></tr>
<tr><td>Solids in slurry discharge (bleed), wt%</td><td colspan="2">12</td></tr>
<tr><td>Solids in sludge to disposal, wt%</td><td colspan="2">80</td></tr>
</table>

Table II

Filtration and Settling Rate Assumptions Used in Sizing Dewatering Equipment

	Unit area requirements		
	Thickener, ft2/ton/day	Clarifier, ft2/ton/day	Filter, lb/hr/ft2
Conventional - base case	2.00	---	200
Auburn - Case I	0.97	2.00	---
Auburn - Case II	0.77	2.00	350

Table III

Process Equipment List and Cost

Conventional Dewatering System - Base Case

Item	No.	Description	Equipment cost, 1982$
1. Thickener and rake	1	108-ft diameter, carbon steel tank and concrete basin with epoxy coatings, rake, and 20-hp motor	$ 700,000
2. Pump, thickener underflow	2	1,152 gpm, 75-ft head, neoprene lined, 50 hp (1 operating and 1 spare)	20,000
3. Filter, rotary drum	4	558 ft^2, 10-ft diameter by 18-ft long drum, stainless steel, including vacuum and filtrate pumps and discharge conveyor (3 operating and 1 spare)	955,000
Total process equipment			$1,675,000

Table IV

Process Equipment List and Cost

Auburn Dewatering System - Case II

Item	No.	Description	Equipment cost, 1982$
1. Thickener and rake	1	75-ft diameter, carbon steel tank and concrete basin with epoxy coatings, rake, and 10-hp motor	$434,000
2. Pump, thickener underflow	2	394 gpm, 75-ft head, neoprene lined, 30 hp (1 operating and 1 spare)	19,000
3. Clarifier (Lamella type)	1	10,080-ft^2 settling area, carbon steel tank and concrete basin with epoxy coatings, fiberglass settling plates, and rake	271,000
4. Pump, clarifier underflow	2	1,152 gpm, 75-ft head, neoprene lined, 50 hp (1 operating and 1 spare)	20,000
Total process equipment			$744,000

Table V

Process Equipment List and Cost

Auburn Dewatering System - Case II

Item	No.	Description	Equipment cost, 1982$
1. Thickener and rake	1	68-ft diameter, carbon steel tank and concrete basin with epoxy coatings, rake, and 10-hp motor	$ 388,000
2. Pump, thickener underflow	2	571 gpm, 75-ft head, neoprene lined, 30 hp (1 operating and 1 spare)	19,000
3. Clarifier (Lamella type)	1	10,080-ft^2 settling area, carbon steel tank and concrete basin with epoxy coatings, fiberglass settling plates, and rake	271,000
4. Pump, clarifier underflow	2	1,152 gpm, 75-ft head, neoprene lined, 50 hp (1 operating and 1 spare)	20,000
5. Filter, rotary drum	3	500 ft^2, 10-ft diameter by 16-ft long drum, stainless steel, including vacuum and filtrate pumps and discharge conveyor (2 operating and 1 spare)	641,000
Total process equipment			$1,339,000

TABLE VI

Total Capital Investment

Conventional Dewatering System - Base Case[a]

	Total, k$	Percent of total capital investment
Process equipment	1,675	27.6
Piping and insulation	117	1.9
Foundation and structural	84	1.4
Excavation, site preparation, and roads	84	1.4
Electrical	419	6.9
Instrumentation	50	0.8
Buildings	206	3.4
Subtotal	2,635	43.4
Services and miscellaneous	211	3.5
Subtotal direct investment	2,846	46.9
Engineering design and supervision	228	3.7
Architect and engineering contractor	85	1.4
Construction expense	512	8.4
Contractor fees	171	2.8
Subtotal	3,842	63.2
Contingency	768	12.6
Subtotal fixed investment	4,610	75.8
Allowance for startup and modifications	369	6.1
Interest during construction	1,097	18.1
Total capital investment	6,076	100.0

---

a. Basis: 1982 dollars; equipment shown in Table III; Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

TABLE VII

Total Capital Investment

Auburn Dewatering System - Case I[a]

	Total, k$	Percent of total capital investment
Process equipment	744	29.9
Piping and insulation	52	2.1
Foundation and structural	37	1.5
Excavation, site preparation, and roads	37	1.5
Electrical	186	7.5
Instrumentation	22	0.9
Buildings	---	---
Subtotal	1,078	43.4
Services and miscellaneous	86	3.5
Subtotal direct investment	1,164	46.9
Engineering design and supervision	93	3.7
Architect and engineering contractor	35	1.4
Construction expense	210	8.4
Contractor fees	70	2.8
Subtotal	1,572	63.2
Contingency	314	12.6
Subtotal fixed investment	1,886	75.8
Allowance for startup and modifications	151	6.1
Interest during construction	449	18.1
Total capital investment	2,486	100.0

---

a. Basis: 1982 dollars; equipment shown in Table IV; Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

TABLE VIII

Total Capital Investment

Auburn Dewatering System - Case II[a]

	Total, k$	Percent of total capital investment
Process equipment	1,339	27.9
Piping and insulation	94	2.0
Foundation and structural	67	1.4
Excavation, site preparation, and roads	67	1.4
Electrical	335	7.0
Instrumentation	40	0.8
Buildings	140	2.9
Subtotal	2,082	43.4
Services and miscellaneous	167	3.5
Subtotal direct investment	2,249	46.9
Engineering design and supervision	180	3.7
Architect and engineering contractor	67	1.4
Construction expense	405	8.4
Contractor fees	135	2.8
Subtotal	3,036	63.2
Contingency	607	12.7
Subtotal fixed investment	3,643	75.9
Allowance for startup and modifications	291	6.1
Interest during construction	867	18.0
Total capital investment	4,801	100.0

---

a. Basis: 1982 dollars; equipment shown in Table V; Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

Table IX

Total Annual Revenue Requirements - Regulated Utility Economics

Conventional Dewatering System - Base Case[a]

Direct Conversion Costs	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Operating labor and supervision	8,320 man-hr	15.00/man-hr	124,800	8.9
Maintenance - plant labor and materials 3% of direct investment			85,400	6.1
Electricity	1,620 kkWh	0.037/kWh	59,900	4.2
Analyses	1,000 man-hr	21.00/man-hr	21,000	1.5
Subtotal direct conversion costs			291,100	20.7
Indirect Costs				
Capital charges - 16.1% of total capital investment			978,200	69.5
Overhead - plant and administrative 60% of conversion costs less utilities			138,700	9.8
Subtotal indirect costs			1,116,900	79.3
Total annual revenue requirements			1,408,000	100.0

---

a. Basis: Midwest plant location; mid-1984 operating costs.
Total capital investment - $6,076,000; total direct investment $2,846,000.
Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

Table X

Total Annual Revenue Requirements - Regulated Utility Economics

Auburn Dewatering System - Base Case I[a]

Direct conversion costs	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Operating labor and supervision	8,320 man-hr	15.00/man-hr	124,800	17.6
Maintenance-plant labor and materials 3% of direct investment			34,900	4.9
Electricity	565 kkWh	0.037/kWh	20,900	2.9
Analyses	1,000 man-hr	21.00/man-hr	21,000	3.0
Subtotal direct conversion costs			201,600	28.4
Indirect costs				
Capital charges - 16.1% of total capital investment			400,200	56.3
Overhead - plant and administrative 60% of conversion costs less utilities			108,400	15.3
Subtotal indirect costs			508,600	71.6
Total annual revenue requirements			710,200	100.0

---

[a] Basis: Midwest plant location; mid-1984 operating costs.
Total capital investment - $2,486,000; total direct investment $1,164,000.
Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

Table XI

Total Annual Revenue Requirements - Regulated Utility Economics

Auburn Dewatering System - Base Case II[a]

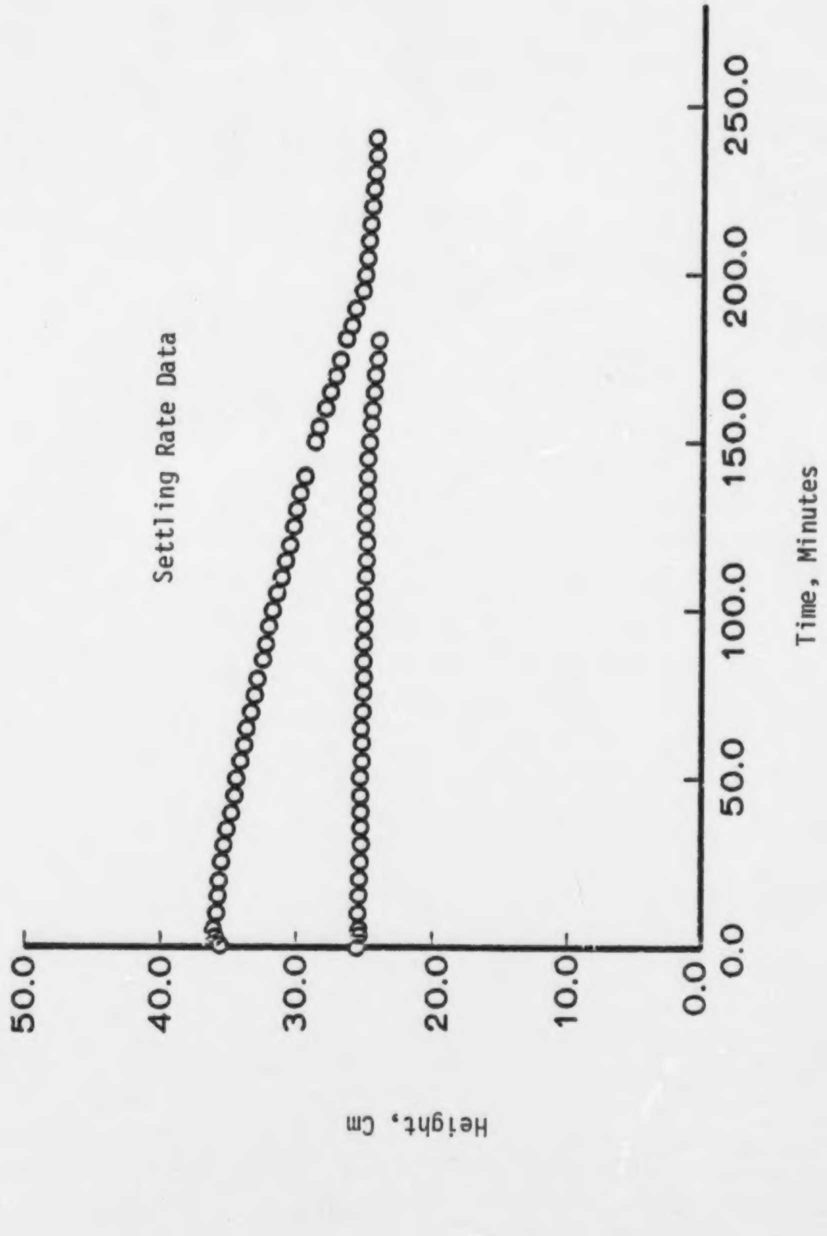
	Annual quantity	Unit cost, $	Total annual revenue requirements, $	Percent of total annual revenue requirements
Direct conversion costs				
Operating labor and supervision	8,320 man-hr	15.00/man-hr	124,800	10.7
Maintenance - plant labor and materials 3% of direct investment			67,500	5.8
Electricity	1,343 kkWh	0.037/kWh	49,700	4.3
Analyses	1,000 man-hr	21.00/man-hr	21,000	1.8
Subtotal direct conversion costs			263,000	22.6
Indirect costs				
Capital charges - 16.1% of total capital investment			773,000	66.4
Overhead - plant and administrative 60% of conversion costs less utilities			128,000	11.0
Subtotal indirect costs			901,000	77.4
Total annual revenue requirements			1,164,000	100.0

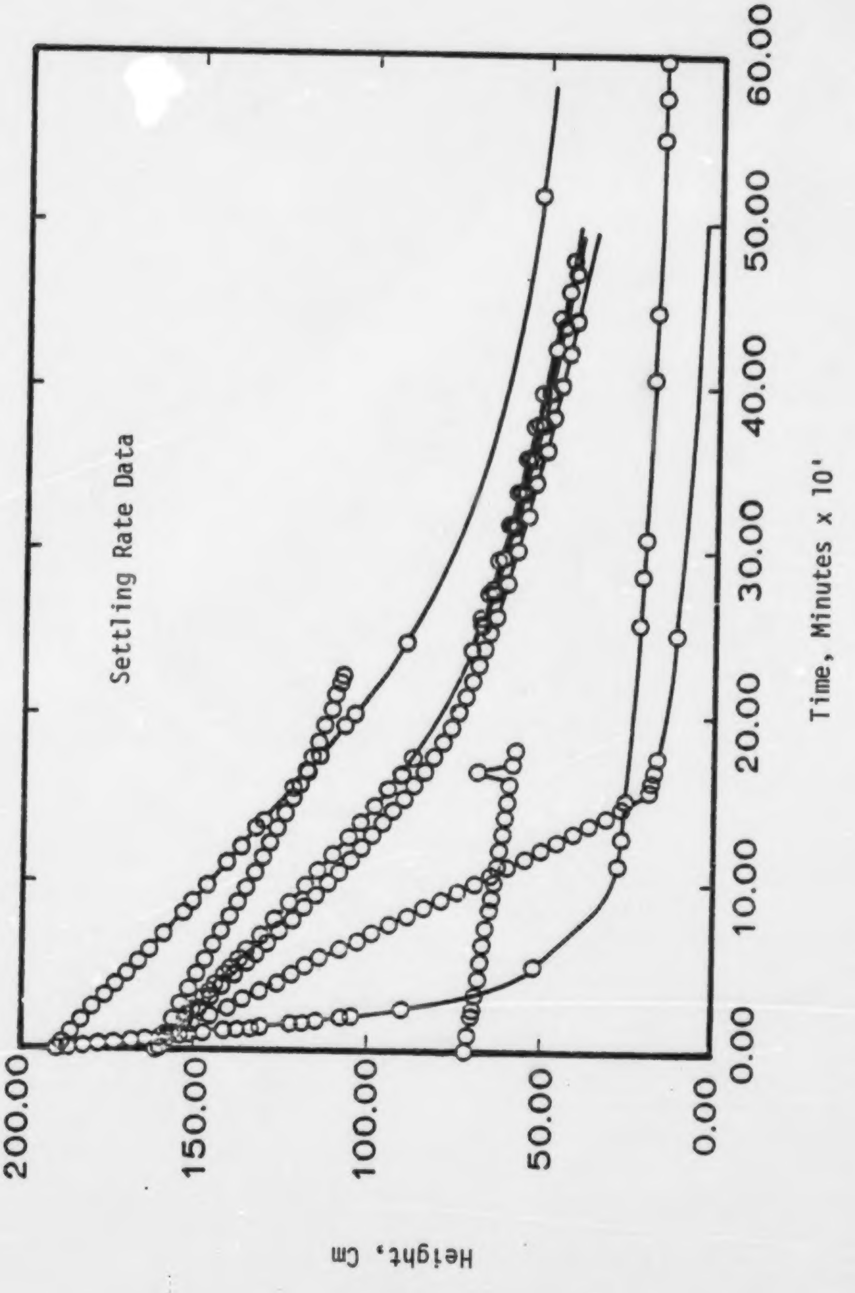
---

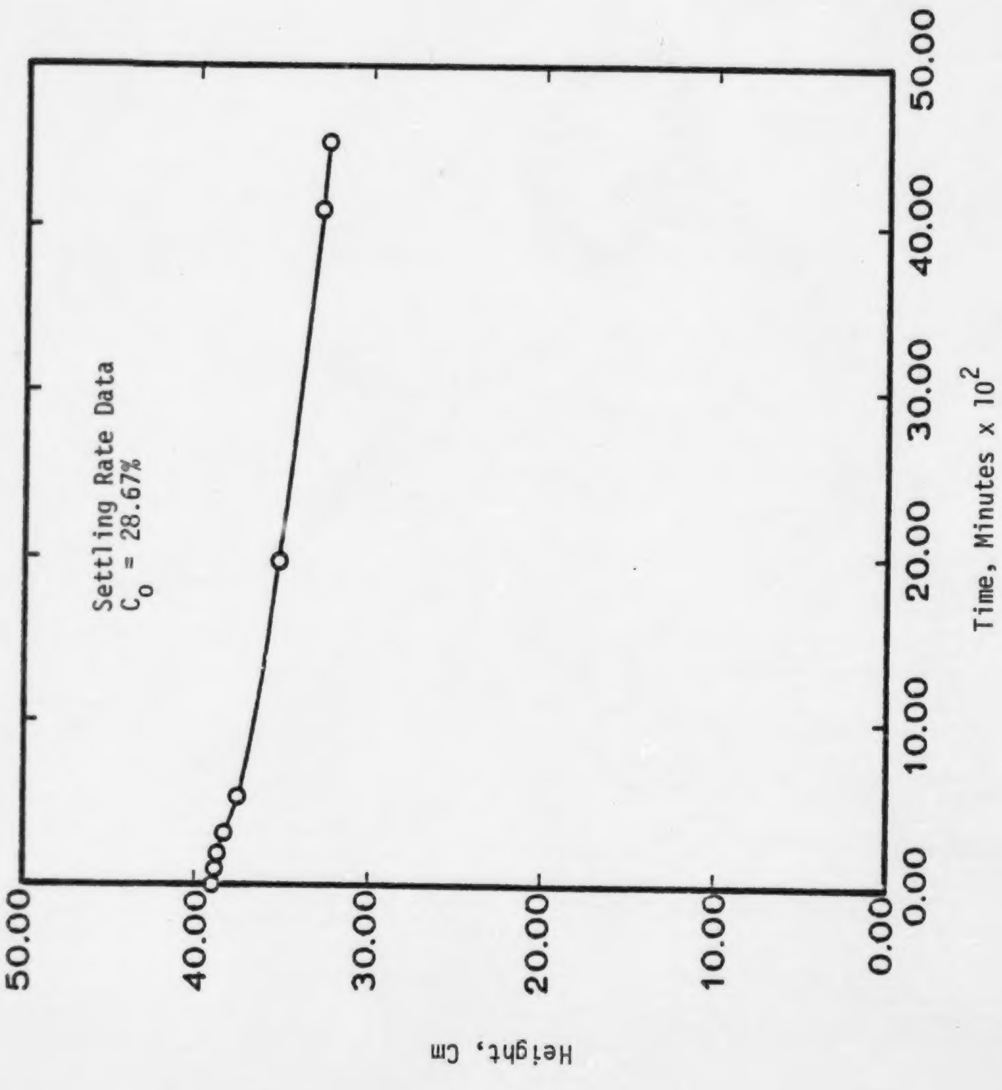
a. Basis: Midwest plant location; mid-1984 operating costs.
Total capital investment - $4,801,000; total direct investment $2,249,000.
Widows Creek units 7 and 8 limestone scrubber design conditions shown in Table I.

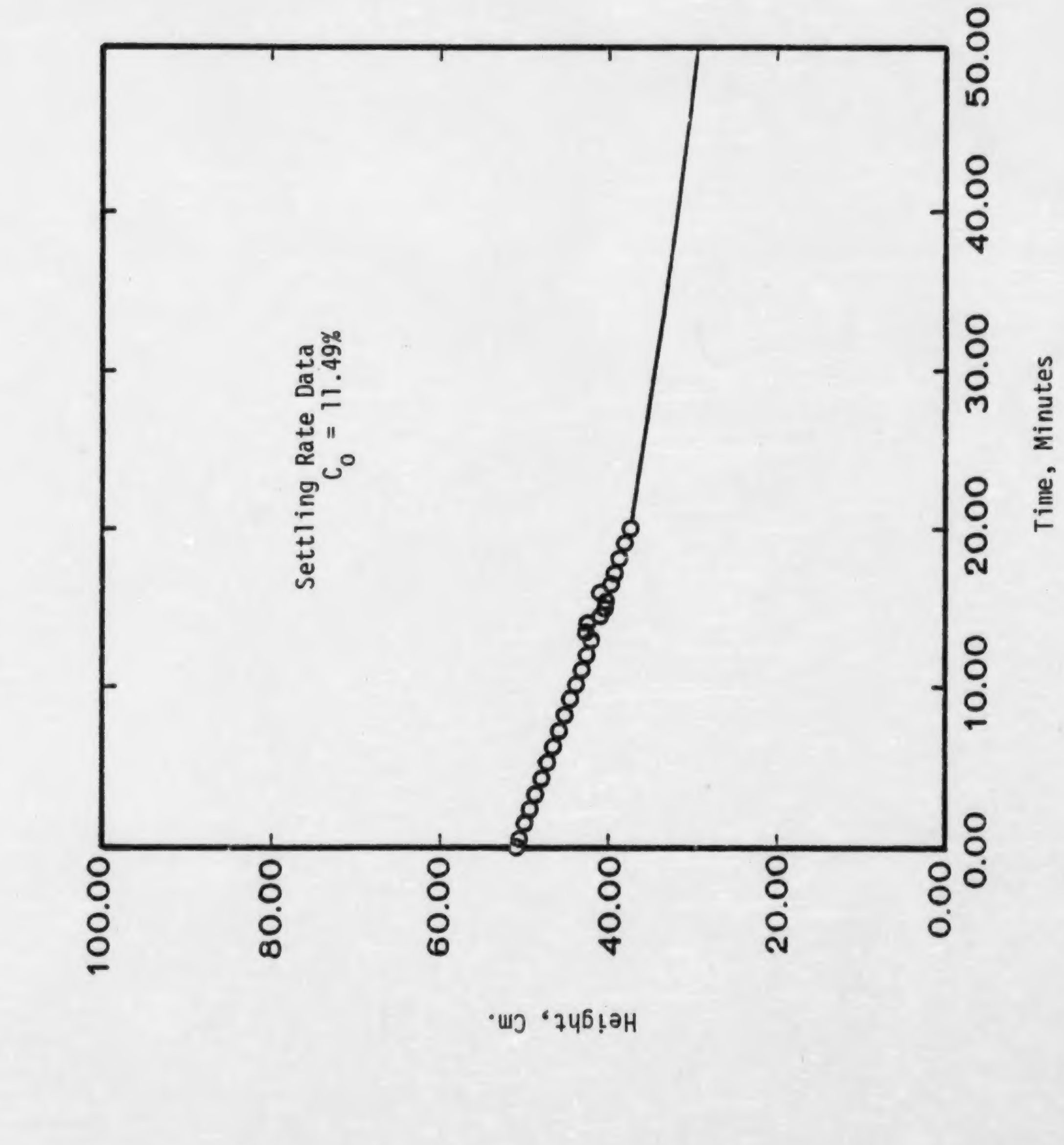
Appendix G

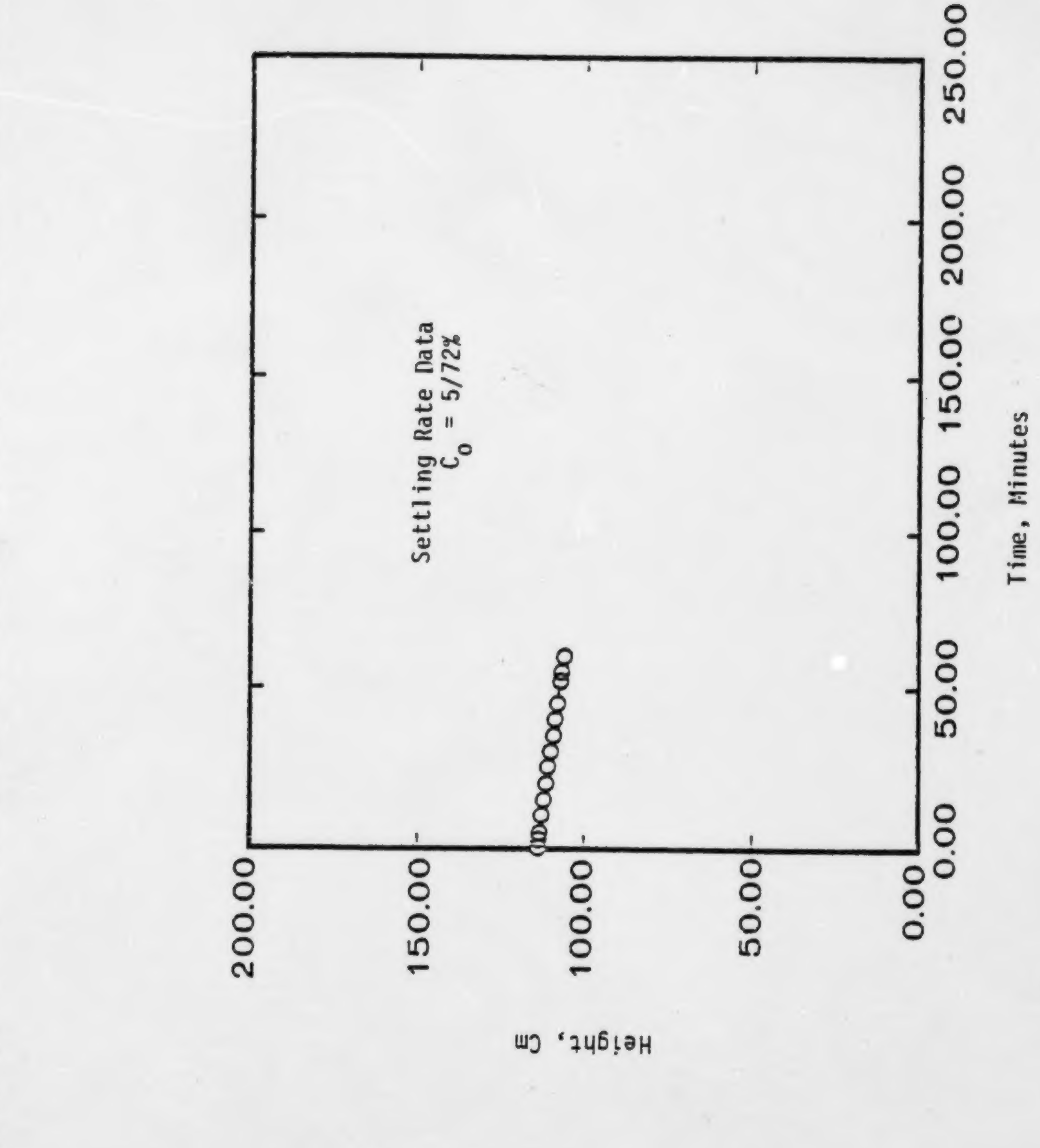
Batch Settling Data: Characterization of FGD Waste Used in Pilot Evaluation


Settling Rate Data
Height, Cm
50.0
40.0
30.0
20.0
10.0
0.0
0.0
50.0
100.0
150.0
200.0
250.0
Time, Minutes


Settling Rate Data
Height, Cm
200.00
150.00
100.00
50.00
0.00
Time, Minutes x 10'
0.00
10.00
20.00
30.00
40.00
50.00
60.00


Settling Rate Data
C_o = 28.67%
Height, Cm
Time, Minutes x 10^2
0.00
10.00
20.00
30.00
40.00
50.00
0.00
10.00
20.00
30.00
40.00
50.00


Settling Rate Data
C_o = 11.49%
Height, Cm.
100.00
80.00
60.00
40.00
20.00
0.00
0.00
10.00
20.00
30.00
40.00
50.00
Time, Minutes


Settling Rate Data
C_o = 5/72%
Height, Cm
200.00
150.00
100.00
50.00
0.00
0.00
50.00
100.00
150.00
200.00
250.00
Time, Minutes